PB-241 310

SYSTEM ALTERNATIVES IN OXYGEN ACTIVATED SLUDGE

John B. Stamberg, et al

District of Columbia Department of Environmental Services

Prepared for:

National Environmental Research Center

April 1975

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ABSTRACT

An oxygen activated sludge system consisting of a unique gas-tight biological reactor, gravity clarification and solids handling equipment was operated on District of Columbia primary effluent during a two-year period over a wide range of loading (F/M 0.26 to 2.0) with Solids Retention Times (SRT) from 2.0 to 3.0 days at the EPA-DC Pilot Plant. The reactor detention times varied from a nominal 1.5 to over 3.0 hours with the mixed liquor suspended solids between 2700 and 8100 mg/l. The clarifier overflow rates varied between 300 and 1960 gpd/ft^2 depending on the mixed liquor particle shape (the intensity of filamentous growth) and water temperature. The average underflow solids varied between 1.0% and 2.4% depending on clarifier volume and recycle rate. The total solids production varied from 0.35 to 1.0 grams of solids produced per gram of BOD_5 applied with as little as 200 pounds of underflow waste per million gallons of flow. The undigested underflow solids thickened to 4-5% by either gravity or air floatation without chemical additives and then dewatered by vacuum to 8-12% with polymers and 13-21% with lime and ferric chloride addition.

This report is submitted in fulfillment of Project 11010 EYM and Contract No. 68-01-0162 by the Department of Environmental Services, Government of the District of Columbia under the sponsorship of the Office of Research and Monitoring, Environmental Protection Agency. Work was completed as of December 1972

FOREWORD

Man and his environment must be protected from the adverse effects of pesticides, radiation, noise and other forms of pollution, and the unwise management of solid waste. Efforts to protect the environment require a focus that recognizes the interplay between the components of our physical environment--air, water, and land. The National Environmental Research Centers provide this multidisciplinary focus through programs engaged in

- studies on the effects of environmental contaminants on man and the biosphere, and
- a search for ways to prevent contamination and to recycle valuable resources.

This report describes new technological advances with pure oxygen in conventional biological treatment of wastewaters. These advances in technology increase treatment reliability and thus reduce environmental contamination.

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Director
National Environmental
Research Center, Cincinnati

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-670/2-75-008	2.	3. PB 241 310
4. TITLE AND SUBTITLE SYSTEM ALTERNATIVES IN OXYGEN ACTIVATED SLUDGE		5. REPORT DATE April 1975; Issuing Date 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) John B. Stamberg, Dolloff F. Bishop, Stephen M. Bennett, and Alan B. Hais		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS EPA-DC Pilot Plant, Department of Environmental Services, Government of the District of Columbia 5000 Overlook Avenue, SW Washington, DC 20032		10. PROGRAM ELEMENT NO. 1BB043 ROAP 21-ASR Task - 015 11. CONTRACT/~~GRANT~~ NO. 68-01-0162
12. SPONSORING AGENCY NAME AND ADDRESS National Environmental Research Center Office of Research and Development U.S. Environmental Protection Agency Cincinnati, Ohio 45268		13. TYPE OF REPORT AND PERIOD COVERED Final Report 14. SPONSORING AGENCY CODE
15. SUPPLEMENTARY NOTES		

16. ABSTRACT

An oxygen activated sludge system with co-current contacting of oxygen and mixed liquor in a plug flow reactor was operated on District of Columbia primary effluent during a two-year period over a wide range of loading (F/M 0.26 to 2.0) with Solids Retention Times (SRT) from 2.0 to 13.0 days at the EPA-DC Pilot Plant. The reactor detention times varied from a nominal 1.5 to over 3.0 hours with the mixed liquor suspended solids between 2700 and 8100 mg/l. The clarifier overflow rates varied between 300 and 1960 gpd/ft^2 depending on the mixed liquor particle shape (the intensity of filamentous growth) and water temperature. The average underflow solids varied between 1.0% and 2.4% depending on clarifier volume and recycle rate. The total solids production varied from 0.35 to 1.0 pound of solids produced per pound of BOD_5. The undigested underflow solids thickened to 4-5% by either gravity or air floatation without chemical additives and dewatered by vacuum filtration to 8-12% with polymers and 13-21% with lime and ferric chloride addition.

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Springfield, VA. 22151

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
*Oxygen Aerobic bacteria Sedimentation Thickening Waste water Activated sludge process Centrifuges Filtration *Settling	*Oxygen activated sludge Suspended solids Sludge dewatering Sludge production	13B

PRICES SUBJECT TO CHANGE

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(This Report)*	21. NO. OF PAGES
RELEASE TO PUBLIC	UNCLASSIFIED	70
	20. SECURITY CLASS *(This page)* UNCLASSIFIED	22. PRICE $4.25-3.25

☆ U.S. GOVERNMENT PRINTING OFFICE: 1975-657-592/5348 Region No. 5-11

EPA-670/2-75-008
April 1975

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Contract No. 68-01-0162
Project 11010 EYM
Program Element 1BB043

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OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

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ACKNOWLEDGEMENTS

The efforts of Dr. Robert Samworth, Director of Research and Development Projects, Department of Environmental Services, Government of the District of Columbia were especially helpful.

On the EPA-DC Pilot Plant staff, Chief Operators George Gray and Herbert Braxton and their crew chiefs Kenneth Graham, Jack Leech, John Norton, Calvin Taylor and Kenneth Taylor were vital to a successful pilot plant operation. The support from the instrument staff headed by Walter Schuk and Paul Ragsdale and the mechanical staff, headed by Robert Hallbrook and Charles Euth were equally important. The efforts of Paul Warner and his analytical staff are greatly appreciated.

The support by the Linde Division of Union Carbide to initiate the study and to work with the pilot plant during the first six months of operation is appreciated.

SECTION I

CONCLUSIONS

1. A gas-tight biological oxygen reactor with independent control of dissolved oxygen and mixing and a gravity clarifier under aerobic conditions produces a good quality secondary effluent from District of Columbia primary effluent in operations with average reactor detention times between 1.5 to 2.5 hours (based on influent flow) and with MLSS concentrations between 4000 and 8000 mg/l.

2. Biodegradable organics in the Washington, D.C. primary effluent are essentially completely insolubilized by the oxygen process (less than 5 mg/l of soluble BOD). Total carbonaceous BOD removal depends upon the amount of suspended solids in the effluent and, therefore, on the clarification efficiency.

3. Oxygen micro-organisms are visually the same as those in a parallel typical conventional step aeration system; however, the rate of activity of the oxygen volatile solids is greater above an SRT of 6 days.

4. As also observed in the air system, oxygen activated sludge is subject to filamentous (Sphaerotilus) growth when operated below an SRT of 5 days on D.C. primary effluent.

5. At SRT's above 6 days, total production of excess biological solids is significantly lower in the EPA-DC oxygen system than in the parallel step aeration system. As little as 0.35 grams of excess solids are produced per gram BOD added at an SRT of 13 days in the oxygen process.

6. Nitrification is achieved in the oxygen aeration system in the summer and fall at EPA-DC Pilot Plant.

7. Based upon the influent and exhaust gas flows, over 95% of the input oxygen is consistently utilized in the EPA-DC oxygen system.

8. Oxygen requirements are based on the COD removal (influent COD minus effluent COD and waste COD) and the nitrogenous oxygen demand. As the SRT increases and endogenous respiration occurs, increased oxygen is required to bio-oxidize the solids. Also at warm temperatures and high SRT's, nitrification occurs with further oxygen demands.

9. Sludge in the oxygen system underflow settles to approximately 1.0-1.4% solids by weight in a clarifier with 1.9 hours of hydraulic retention time and 2.0% to 2.4% in a clarifier with 2.8 hours of hydraulic retention time.

10. When the oxygen clarifier is operated with a deep feed-well or with the mixed liquor sufficiently concentrated to settle in a subsidence (zone) settling pattern, the blanket acts as a filter and produces high quality effluent. In the summer and fall with zone settling, the deep well clarifier operates at a peak rate of 1940 gpd/ft^2. A larger clarification area is required in the winter than in the summer on District of Columbia wastewater for a given MLSS concentration.

11. With a shallow center feed-well and with the mixed liquor concentrations low enough (under 4000 mg/l) to permit discrete particle settling, better settling rates are observed in the clarifier than with the sludge blanket method of clarifier operation (described in No. 12 above). At Washington only moderate decreases in effluent quality (increase in SS from 15 to 25 mg/l) are observed with the shallow center well operation.

12. With the normally high mixed liquor concentrations used in the oxygen aeration process, design of clarifiers should consider overflow rates, volume, and solids loading.

13. In general, air floatation clarification is not a suitable alternative to gravity clarification because, even though the concentration of the recycled solids are increased, the effluent suspended solids are much higher than in a comparably loaded gravity clarifier.

14. In Washington, D.C., non-bulking (SVI less than 200) oxygen activated sludge settling characteristics are essentially similar to non-bulking air activated sludges, nitrifying sludges and denitrifying sludges. Likewise, bulking oxygen activated sludges settle similarly to bulking air sludges.

15. Excess waste oxygen solids thicken to 4.5% solids by either gravity or air floatation thickening.

16. Waste oxygen solids and primary sludge mixed at a 0.4:1 weight ratio of secondary (oxygen) to primary solids gravity thickens up to 11.0% solids.

17. Excess waste oxygen solids alone dewater by vacuum filtration to 8-12% solids with polymer conditioning and to 13-21% with $FeCl_3$ and lime conditioning. In a filter press with $FeCl_3$ and lime conditioning, these solids dewater to 20.6% solids.

18. Proper coordination of the reactor, clarifier and solids handling facilities permit many alternative designs of oxygen activated sludge. In general, as the reactor size is increased, a lower MLSS concentration is required for a given biological state (F/M ratio). The lower the MLSS concentration, the smaller the required secondary clarification area.

19. The solids handling requirement of an oxygen system depend on the biology established in the reactor/clarifier combination. If minimum excess biological sludge production is required, then more capacity is required in the reactor/clarifier combination and in the oxygen supply. Further, the concentration of the clarifier underflow solids is dependent on clarifier volume and recycle rate.

SECTION II

RECOMMENDATIONS

1. The oxygen activated sludge system fed with raw D.C. wastewater should be evaluated over a wide range of operation conditions.

2. The oxygen activated sludge system, fed either raw or primary wastewater, should be evaluated as a high rate system with high F/M ratios and very low SRT's.

3. The use of alum precipitation of phosphorus with chemicals such as lime to restore pH and alkalinity should be evaluated.

4. Other methods of process operation such as step aeration and contact stabilization should be evaluated with the oxygen activated sludge process.

5. The use of oxygen digestion of primary, secondary or a combination of waste sludges in a separate digester should be investigated as a potential system alternative to operation of the oxygen reactor at low F/M.

6. A comprehensive thickening-dewatering study is required on a variety of equipment to evaluate the handling of the waste oxygen solids.

7. Liquid solids separation methods as alternates to gravity clarification should be considered.

8. A method of designing full scale circular and rectangular clarifiers from continuous pilot plant and batch test information is badly needed not only for oxygen but for all biological sludges.

9. Separate nitrification utilizing oxygen should be evaluated.

10. An accurate method for measuring the bulk particle density of biological solids is needed for defining settling characteristics.

11. A study using daily settling data should be undertaken to observe the effect of grease accumulation and/or inerts (i.e. silt and clay) on bulk particle density and on the corresponding settling velocity.

SECTION III

INTRODUCTION

The use of pure oxygen within the activated sludge process dates back some 20 years. Pirnie (1) proposed a system of predissolving pure oxygen in high concentrations in the influent wastewater before entering a non-aerated mixed reactor. The process was termed "a bio precipitation system." Biological success was achieved by Okun (2) in bench scale tests and later by Budd and Lambeth (3) on a pilot scale, but oxygen utilization efficiencies of 20-25% were too costly. Okun and Lynn (4) and later Okun (5) showed an increase in the effective sludge activity in the mixed liquor by reducing or eliminating anaerobic periods such as can occur in clarification. McKinney and Pfeffer (6) more recently reviewed the use of oxygen in activated sludge. Increased metabolism rates, produced by eliminating periods of zero dissolved oxygen, would increase treatment efficiencies in overloaded plants and reduce the size required for new plants. Thus, potential reductions in capital investment were viewed possible for oxygen systems.

Union Carbide recently developed the UNOX System (7) which is an oxygen-aeration-activated sludge system with an oxygen utilization of over 90%. This oxygen-activated sludge process (UNOX) is presently being piloted in several locations. The Environmental Protection Agency participated in the earlier studies in Batavia, New York and St. Louis, Missouri and presently is participating in studies at Las Virgenes, California, Newtown Creek, New York City, New York and Washington, D.C. (8). In Washington, D.C. the EPA-DC Pilot Plant has been evaluating the oxygen-activated sludge process since May of 1970, nearly two and a half years. This is the longest running pilot study and the only facility that has operated for more than a full year at the same location.

The objectives of the study are to determine process performance and operating requirements on D.C. primary wastewater over a wide range of seasonal diurnal and operational conditions. The oxygen activated sludge process should be and was viewed and evaluated as a system with three interrelated subsystems, the oxygen biological reactor, the clarifiers and the solids handling system.

SECTION IV

ANALYTICAL PROCEDURES

To evaluate the oxygen-activated sludge system, appropriate samples were manually composited over a 24-hour period. Samples were stored at 3°C to minimize biological activity.

The five day biological oxygen demand (BOD) of the composite samples was determined by the probe method (9); the ammonia (9) and nitrate-nitrite (10) on a Technicon Automatic Analyzer. The total organic carbon (TOC) (11) was measured on a Beckman Carbonaceous Analyzer. The total phosphorus (12) was determined by the persulfate method. All other analyses employed Standard Methods (13). Soluble phosphorus and soluble BOD were filtered through a standard glass suspended solids filter before analyses.

SECTION V

BIOLOGICAL REACTOR PERFORMANCE OF OXYGEN ACTIVATED SLUDGE

The first and most unique aspect of the system is the gas tight biological reactor (Figure 1). In the EPA-DC Pilot Plant, primary effluent from the District of Columbia's plant is fed to the oxygen reactor either on steady state flow or on a predetermined daily cycle (diurnal variation), normally with a 2.3:1 (45-105 gpm) daily flow variation but occasionally with an alternate variation of 2.0:1 (40-80 gpm).

With all four available stages, the 8,100-gallon EPA-DC oxygen reactor provides 1.95 hours or detention time at the nominal influent flow of 100,000 gpd. At the peak daily flow, the detention time is 1.29 hours. With three of the available four stages, the detention times are reduced to 1.50 hours and 1.00 hours, respectively, at the nominal and normal peak daily flows.

The reactor is sealed to prevent loss of oxygen and includes submerged hydraulic entrances and exits as well as simple water-sealed mixing equipment. Internal spray equipment using tap water is provided to suppress foam. A partially submerged baffle plate before the internal exit trough retains the foam until the baffle plate is raised to allow the foam build-up to escape. The reactor is staged to provide the proper tank geometry for efficient mixing and oxygen usage.

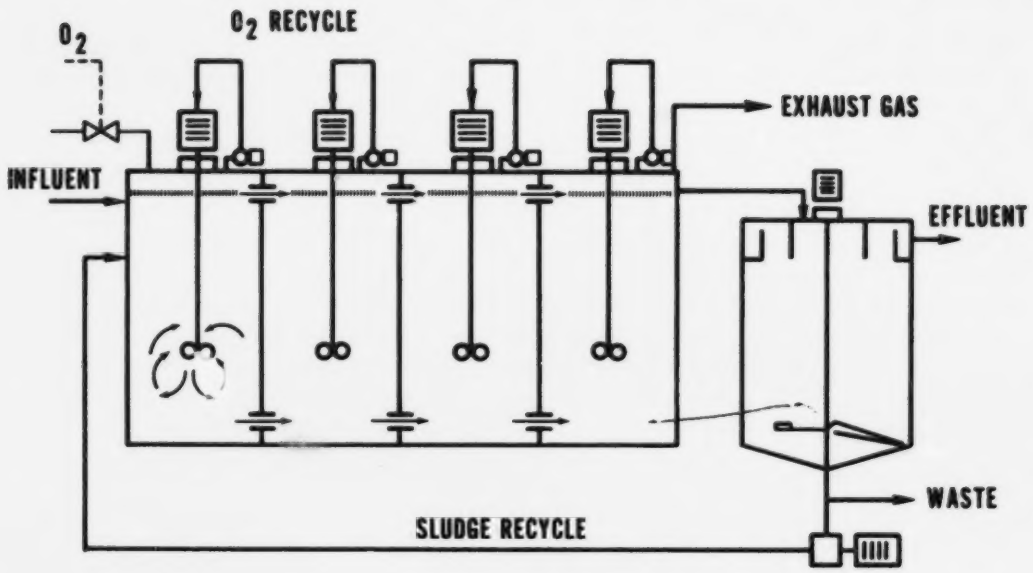

Figure 1. Oxygen Activated Sludge System

Efficient oxygen usage is achieved by co-current contacting of the mixed liquor and oxygen gas through the various stages. The addition of pure oxygen to the reactor is controlled by a pressure regulator. An inlet oxygen control valve actuated by a pressure regulator maintains the overhead gas at a selected pressure usually between 1" and 4" of water. Even with large instantaneous fluctuations in oxygen consumption, the oxygen control valve maintains the selected pressure. The overhead gas pressure is normally selected to produce an oxygen concentration of approximately 50% in the exhaust gas from the last reactor stage. The oxygen is introduced to the first stage where the peak oxygen demand occurs. As the oxygen is used in biological metabolism, respirated carbon dioxide and stripped inert gases reduce the oxygen concentration in the overhead gas flowing co-currently with the mixed liquor through the succeeding stages. The successive decrease of both oxygen availability and oxygen demand produces efficient oxygen use before the residual gas is exhausted from the reactor.

Mixed liquor dissolved oxygen levels in the Blue Plains oxygen reactor are held between 4.0 and 8.0 mg/l by adjusting the recirculation rate of the oxygen gas within the individual stages. The compressor in each stage pumps the overhead gas through the rotating submerged turbine-sparger to provide efficient dispersion and mixing of the recirculated gas. The recirculation rate in each stage may be set either manually on the basis of the dissolved oxygen analysis or automatically using a control system with a dissolved oxygen sensor. The gas recirculation rate in the first stage typically is 3-7 cfm and 1-2 cfm in each of the last three stages. Total recirculation requirements vary between 0.10 and 0.20 ft^3/gal. of flow.

With its high oxygen transfer capabilities (which are essentially independent of turbine mixing rates), the oxygen system is able to operate at high MLSS concentrations. These factors enable the system to readily adsorb shock organic loads. Also, toxic shock loads are better handled, much as in a totally mixed activated sludge system. Both types of systems initially expose the toxic substrate to a large mass of active solids and the resulting "biological inertia" buffers the toxicity.

The oxygen mixed liquor at the EPA-DC Pilot Plant was similar visually to the micro-organisms in conventional activated sludge (Figure 2). The mixed liquor biota was normally very well bio-flocculated with active stalked cilates growing on the bacterial mass. Zooflagellates and free-swimming cilates, although few in number, remained adjacent to or within the flocculated particles. Several varieties of large active rotifers were present in abundance. A few nematodes existed in the sludge. Normally, filamentous growth was not apparent. There was almost complete absence of fragmented debris or unflocculated bacteria between the discrete particles.

Figure 2. Photomicrograph of Oxygen Mixed Liquor

In the sludge retention time (SRT) range below 5 days both the oxygen activated sludge and the conventional aeration systems exhibited filamentous growth on the District of Columbia wastewater. (SRT is the inverse of growth yield). Filamentous growth did not occur during operation above an SRT of 5 days. Normally, when encountering filamentous growth for a few days, reducing system influent flow to increase the SRT reestablished a filamentous free sludge in several days. However, after extended periods of operation with filamentous growth, the Sphaerotilis became firmly entrenched and could not be quickly purged from the system by flow reduction techniques. Hydrogen peroxide added to the recycle in two 24-hour periods approximately a week apart at dosage of 200 mg/l (based on influent flow) was then required to purge the system of filamentous growth.

Pertinent reactor variables and operating conditions are summarized in Table 1 for approximately 2-1/2 years of operation on the EPA-DC oxygen aeration pilot plant. The food to micro-organism ratio (F/M) varied from 2.0 to 0.27 grams BOD_5 applied/gram of MLVSS. The Solids Retention Time (SRT) varied from less than 2 days to 13 days. The average volumetric loadings were as high as 185 lb BOD_5 applied 1,000 ft^3 of tank.

The reduction of BOD in the reactor was excellent. With an influent BOD up 130 mg/l, a wide range of detention times from 1.5 to 2.5 hours and SRT that varied from 13 to as low as 2 days, the effluent soluble BOD was consistently less than 5 mg/l as described in Table 2. Virtually complete insolubilization of the BOD in the primary effluent occurred and, thus, BOD removal on the EPA-DC oxygen system was a function of clarification efficiency.

On the District of Columbia wastewater, as seen in Figure 3, the volatile portion of the oxygen solids exhibited a much higher activity for F/M ratios in the SRT range above 6 days than the EPA-DC's step aeration pilot process. Figure 3 indicates that a lower total volatile mass under aeration was required with oxygen than with air to obtain any given SRT above 6 days for a similar influent BOD. Thus, shorter detention times were possible with oxygen than with step aeration for similar MLSS concentrations to achieve any given SRT above 6 days. Further, at identical SRT's above 6 days, the oxygen system produced less excess biological solids. The most probable reason for the increased activity was attributed to maintaining the mixed liquor dissolved oxygen between 4 and 8 mg/l. The independently controlled mixing also minimized sludge pockets, dead spots, and shearing of the floc particles. Mixed liquor entering the clarifier had a high dissolved oxygen content which permitted a certain amount of aerobic metabolism in the clarifier and greatly reduced the time that the bio-mass was in an anaerobic condition.

TABLE 1

REACTOR VARIABLES AND OPERATING CONDITIONS
EPA-DC PILOT PLANT
1970

	Flow Rate gpm	Aeration Time (hr)	Recycle Rate	MLSS mg/l	MLVSS %	SRT days	F/M (lb BOD/day/ lb MLVSS	Volumetric loading (lb BOD/day/1000 ft3)	Mixer Power k.w. - hr / 1,000 gal	Compresser Power k.w.-hr / 1,000 gal	Temp. °F
June 12-30	50-55	2.00	50%	4140	74%	7.7	0.333	57	1.27	0.39	74-80
July	80	1.66	50%	5180	70%	7.3	0.342	80	1.19	0.28	78-84
August 1-25	80	1.66	42%	5250	73%	11.8	0.296	80	0.98	0.39	82-85
September	70*	1.95	32%	6000	78%	10.7	0.304	96	0.92	0.41	79-83
October 3-11	70*,**	1.95	38%	8120	67%	5.5	0.283	106	1.00	0.35	70-79
November 10-30	70*	1.95	37%	6350	73%	5.5	0.355	108	1.00	0.26	66-69
January 1-16	53	2.50	77%	5300	80%	13.0	0.275	89	1.18	0.34	58-60
January 17-31	53	2.50	80%	3940	81%	4.7	0.392	80	1.42	0.32	58-60

TABLE 1 (CONTINUED)
REACTOR VARIABLES AND OPERATING CONDITIONS
EPA-DC PILOT PLANT
1971

	Flow Rate gpm	Aeration Time (hr)	Recycle Rate	MLSS mg/l	MLVSS %	SRT days	F/M (lb BOD/day/ lb MLVSS	Volumetric loading (lb BOD/day/1000 ft^3)	Mixer Power k.w.-hr. / 1,000 gal.	Compresser Power k.w.-hr. / 1,000	Temp. °F
March 1-8	60-70	2.15	50-60%	3070	77%	3.7	0.580	90	-	-	60-62
April	31-67	3.30-1.55	90-60%	2710	81%	1.3-4.0	0.30-1.00	98	-	-	62-65
May	60	1.70	65%	2750	78%	2.0	0.970	157	-	-	65-71
June	30-70	3.70-1.50	100-50%	4000	73%	13.0	0.400	95	-	-	70-77
July	70*	1.5	50%	6600	70%	12.6	0.430	160	-	-	[illegible]
August	70*	1.5	50%	7500	70%	10.0	0.32	131	-	-	76-81
September	70*	1.5	46%	7400	72%	7.5	0.39	185	-	-	76-81
October	70*	1.5	36%	6000	73%	9.5	0.31	146	-	-	76-71
November	70*	2.0	30%	4600	78%	9.8	0.39	111	-	-	71-65
December 1-21	70*	2.0	25%	4400	80%	9.0	0.40	111	-	-	65-63
December 22-31	70*	2.0	25%	4200	81%	6.5	0.50	122	-	-	63-61

TABLE 1 (CONTINUED)
REACTOR VARIABLES AND OPERATING CONDITIONS
EPA-DC PILOT PLANT
1972

	Flow Rate gpm	Aeration Time (hr)	Recycle Rate	MLSS mg/l	MLVSS %	SRT days	F/M (lb BOD/day/ lb MLVSS	Volumetric loading (lb BOD/day/1000 ft^3)	Mixer Power $\frac{k.w.-hr.}{1,000\ gal.}$	Compresser Power $\frac{k.w.-hr.}{1,000}$	Temp. °F
January	50	2.8	25%	4300	79%	5.5	0.30	67	-	-	57-63
February	55	2.4	30%	4470	77%	7.3	0.30	62	-	-	54-61
March 1-17	55	2.4	36%	4150	77%	5.1	0.36	70	-	-	58-61
March 21-31	55	2.4	45%	4450	77%	7.3	0.33	70	-	-	59-61
April	55	2.4	47%	4150	75%	7.7	0.38	72	-	-	60-63
May	52	2.6	50%	3700	76%	8.2	0.36	63	-	-	63-71
June	55	2.4	45%	4100	72%	13.0	0.32	60	-	-	70-74

TABLE 2

ORGANIC REMOVAL-EPA-DC PILOT PLANT
1970

	Primary Eff. BOD (mg/l)	Final Eff. BOD (mg/l)	Final Eff. Soluble BOD (mg/l)	Primary Eff. COD (mg/l)	Final Eff. COD (mg/l)	Primary Eff. TOC (mg/l)	Final Eff. TOC (mg/l)	Primary Eff. Suspended Solids (mg/l)	Final Eff. Suspended Solids (mg/l)
June 12-30	89	18	--	250	45	75	14	113	36
July	87	19	--	244	70	65	24	101	53
August 1-25	89	12	2	245	49	77	15	102	28
September	106	13	2	252	51	100	15	107	24
October 3-11	116	14	3	284	51	106	15	120	35
November 10-30	131	27	3	275	63	91	21	92	56
January 1-16	124	11	3	250	59	83	21	98	24
January 17-31	134	32	3	256	99	87	26	100	58

TABLE 2 (CONTINUED)

ORGANIC REMOVAL-EPA-DC PILOT PLANT
1971

	Primary Eff. BOD (mg/l)	Final Eff. BOD (mg/l)	Final Eff. Soluble BOD (mg/l)	Primary Eff. COD (mg/l)	Final Eff. COD (mg/l)	Primary Eff. TOC (mg/l)	Final Eff. TOC (mg/l)	Primary Eff. Suspended Solids (mg/l)	Final Eff. Suspended Solics (mg/l)
March 1-18	121	27	--	251	76	88	22	104	49
April	107	10	4	267	48	81	17	83	18
May	140	7	4	278	51	92	18	120	12
June	110	8	--	238	45	74	18	100	13
July	129	14	--	235	35	78	14	103	11
August	110	15	--	219	32	69	13	97	16
September	149	15	4	239	35	79	15	95	15
October	120	14	4	224	37	69	14	81	15
November	125	20	5	244	54	75	17	90	23
December 1-21	125	20	5	238	59	76	19	95	23
December 22-31	135	18	5	236	53	91	19	95	18

TABLE 2 (CONTINUED)

ORGANIC REMOVAL-EPA-DC PILOT PLANT
1972

	Primary Eff. BOD (mg/l)	Final Eff. BOD (mg/l)	Final Eff. Soluble BOD (mg/l)	Primary Eff. COD (mg/l)	Final Eff. COD (mg/l)	Primary Eff. TOC (mg/l)	Final Eff. TOC (mg/l)	Primary Eff. Suspended Solids (mg/l)	Final Eff. Suspended Solids (mg/l)
January 1-19	121	13	3.5	260	47	78	15	99	20
February	101	15	2.5	224	55	73	21	144	36
March 1-17	113	21	2.5	238	82	79	24	98	46
March 21-31	112	13	3.5	231	56	79	19	118	25
April	116	14	2.5	235	49	75	16	105	22
May	109	20	3.0	229	48	74	18	97	23
June	94	15	3.0	220	43	68	13	96	15

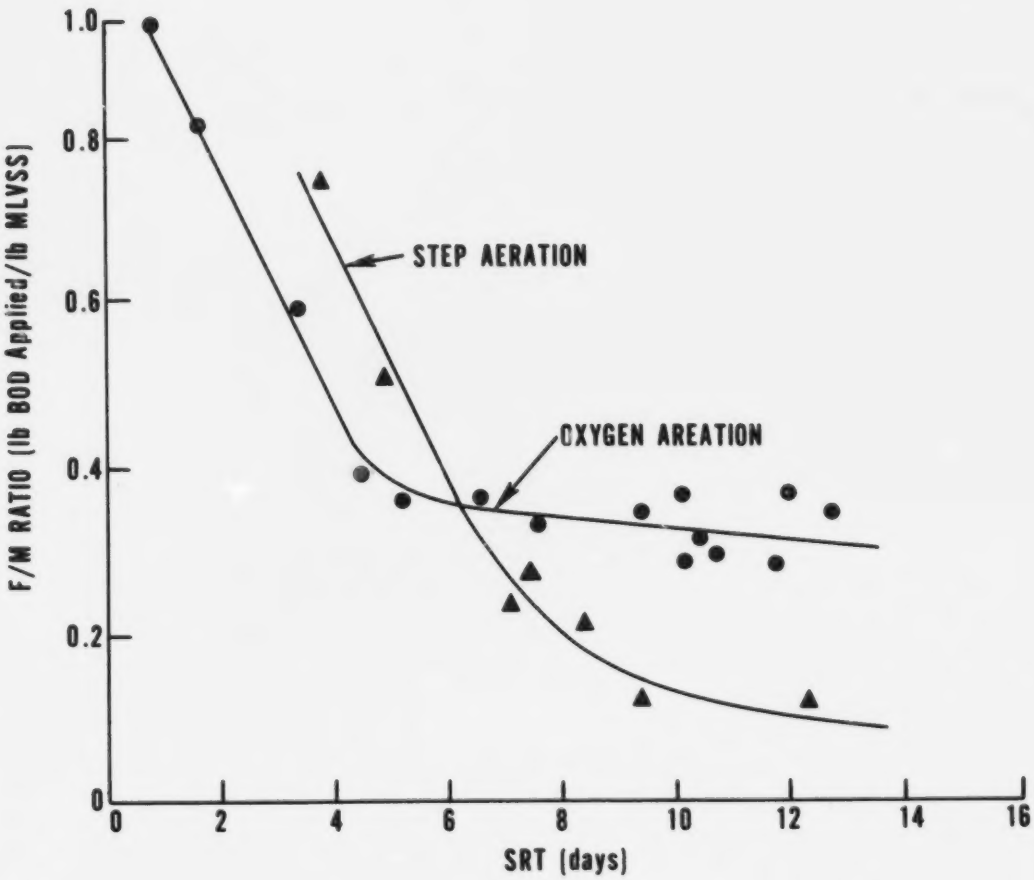

Figure 3. Biological Activity Relationships

The total production of solids in the oxygen system (Figure 4) per pound of BOD added, including underflow waste and effluent solids, was inversely related to the solids retention time (SRT) above an SRT of 1.3 days. The solids production with the oxygen aeration system was significantly lower than in the conventional step aeration pilot system above an SRT of 6 days (or than in a high rate modified air aeration pilot process tested by the EPA-DC staff). Indeed, the total solids production decreased from 0.65 pounds of excess solids per pound of BOD added at an SRT of 6 days to 0.35 pounds of excess solids per pound of BOD added at an SRT of 13 days with only a 33% increase in volatile solids concentration at the higher SRT. The parallel conventional system, operated as step aeration or contact stabilization, exhibited increased solids production through an SRT of 9.5 days with a peak solids production of approximately 1 pound of excess solids per pound of BOD added. However, an approximate four-fold increase in volatile solids was required to raise the SRT from 6 to 13 days in this system and to begin to achieve reduced solids production.

The recirculation of respirated carbon dioxide within the oxygen reactor stages lowered the wastewater pH from 7.0-6.8 in the first stage to 6.4-6.1 in the final stage. With an average system pH of approximately 6.5, the oxygen process more slowly established a nitrifying population than the step aeration activated sludge system operated at a pH of 7.0 to 7.4. However, during the warmer months when the solids wasting was reduced to a level where the nitrifying organisms propagate faster than they were removed, the Nitrosomonas and Nitrobacter populations increased and substantial nitrification occurred in the oxygen system. Nitrogen removal across the oxygen system during periods of high nitrification and partial denitrification was as high as 35-43%. At wastewater temperatures of about 63°F, 5 mg/l of NO_3^--N was still produced with an SRT of 9.0 days. Thus, as in the parallel step aeration process, nitrification in the oxygen system began to decrease in the Fall and became negligible during the Winter. Nitrogen removal decreased to a low of 0-10% during periods without nitrification (Table 3).

Without physical-chemical phosphorus removal, phosphorus was removed from the oxygen system through metabolic uptake and by wasting of excess solids; thus, the removals varied with the metabolism of the mixed liquor. At high SRT's (highly endogenous metabolism), total phosphorus removal averaged only about 15%. At lower SRT's with less endogenous respiration, phosphorus removals increased to 20%.

With alum addition, phosphorus removal in the oxygen system increased as the alum weight ratio (Al^{+++}/P) increased. During experiments conducted in the Fall of 1970, for a dosage equal to an Al^{+++}/P ratio of 1.4/1, 80% of the phosphorus was removed to an average residual

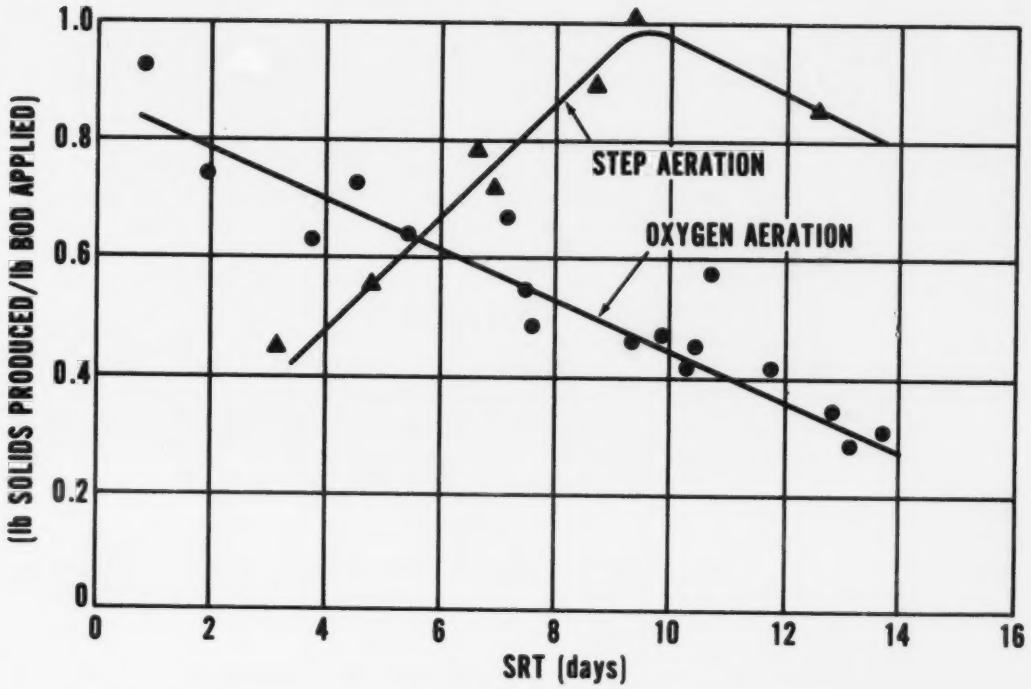

Figure 4. Excess Sludge Production

TABLE 9
NITROGEN SUMMARY - EPA-DC PILOT PLANT
1970

	Primary Effluent TKN lb/mil gal	Primary Effluent (NO_3+NO_2)-N lb/mil gal	Total in lb/mil gal	Effluent TKN lb/mil gal	Final Effluent (NO_3+NO_2)-N lb/mil gal	Waste Solids TKN lb/mil gal	Total Out lb/mil gal
June 12-30	198	0	198	140	3	13	156
July	188	0	188	134	15	4	153
August 1-25	193	0	193	85	28	11	124
September	206	0	206	94	35	19	148
October 3-21	225	0	225	128	22	48	198
November 10-30	212	0	212	166	2	18	186
January 1-16	215	0	215	145	4	17	166
January 17-31	227	0	227	150	3	22	175

TABLE 3
NITROGEN SUMMARY - EPA-DC PILOT PLANT
1971

	Primary Effluent TKN lb/mil gal	Primary Effluent (NO_3+NO_2)-N lb/mil gal	Total in lb/mil gal	Effluent TKN lb/mil gal	Final Effluent (NO_3+NO_2)-N lb/mil gal	Waste Solids TKN lb/mil gal	Total Out lb/mil gal
March 1-18	222	0	222	163	9	12	184
April	212	0	212	138	2	--	140
May	235	0	235	157	0	--	157
June	204	0	204	137	2	--	139
July	187	0	187	78	33	10	121
August	167	0	167	35	59	16	110
September	214	0	214	40	63	19	122
October	196	0	196	60	47	31	138
November	225	1	225	84	52	18	154
December 1-21	215	1	216	94	41	16	151
December 22-31	224	2	226	147	16	25	188

TABLE 3
NITROGEN SUMMARY - EPA-DC PILOT PLANT
1972

	Primary Effluent TKN lb/mil gal	Primary Effluent (NO_3+NO_2)-N lb/mil gal	Total in lb/mil gal	Effluent TKN lb/mil gal	Final Effluent (NO_3+NO_2)-N lb/mil gal	Waste Solids TKN lb/mil gal	Total Out lb/mil gal
January 1-19	196	1	197	127	12	40	179
February	195	4	199	127	6	17	150
March 1-19	203	2	205	144	6	26	176
March 21-31	216	1	217	121	10	30	161
April	200	2	202	113	14	44	171
May	190	1	191	44	57	26	127
June	184	2	186	28	80	10	118

of 1.8 mg/l as P and only a slight decrease in wastewater alkalinity and pH occurred. The filtered effluent (though 0.45 μ) contained an average of 1.6 mg/l of soluble P. When the dosage was increased to a weight ratio of 1.85/1 (Al^{+++}/P), the residual total and soluble phosphorus decreased to 0.62 and 0.53 mg/l as P, respectively. At this higher dosage, however, the buffering capacity of the oxygen mixed liquor was further reduced and the average pH decreased from 6.5 to 6.0. The oxygen bio-mass dispersed and required termination of the alum addition to allow the mixed liquor to recover.

In the low alkalinity D.C. wastewater, additional alkalinity in the form of lime or caustic is required to control pH at a level which will prevent floc dispersion. This pH adjustment may be necessary in either air or oxygen systems but is more likely in an oxygen system because of the increased dissolved CO_2 content of the mixed liquor. The addition of alum and precipitation of $Al(PO_4)$ and $Al(OH)_3$ increases the inert solids carried in the system and adequate clarification for the higher solids concentration must be provided.

Consistently throughout the operation, vented gas from the fourth system reactor stage contained less than 10% of the input oxygen volume. The vented stream was roughly 50% oxygen. Based upon the influent and exhausted oxygen concentrations, the net utilization of oxygen in the process was about 95%. The accountable oxygen consumption consisting of COD removal, nitrification demand, exhaust gas, and effluent dissolved oxygen is summarized in Table 4. The COD removed was calculated by substracting the COD in the underflow waste solids and in the process effluent from the primary effluent COD. With increasing SRT, additional oxygen was required for endogenous respiration (Figure 5). Likewise, during periods with nitrification, additional oxygen was required (Figure 6).

TABLE 4

OXYGEN USAGE-EPA-DC PILOT PLANT
1970

	Primary Eff. COD (lb/mil.gal)	Final Eff. COD (lb/mil.gal)	Waste Sludge COD (lb/mil.gal)	COD Removal from System (lb/mil.gal)	Nitrate Nit. Demand (lb/mil.gal)	Exhaust Oxygen (lb/mil.gal)	Final Eff. D.O. (lb/mil.gal)	Total	Oxygen Supplied (lb/mil.gal)
June 12-30	2080	375	188+	1517	14	85	10*	1626	1750
July	2030	584	42+	1404	69	54	10*	1537	1825
August	2040	408	150+	1482	128	75	10	1695	1775
September	2090	430	217+	1443	160	85*	10	1698	1900
October 3-21	2370	425	630+	1315	100	87	10	1512	1650
November 10-30	2290	525	247	1518	18	85*	10*	1631	±
January 1-16	2080	488	233	1359	9	65	25	1458	1800**
January 17-31	2140	826	336	978	14	65	25	1082	1800**

\+ COD = 1.4 volatile solids
* Estimate
± Inlet meter malfunctioned
** Increase sampling and greater losses of O_2 through sample ports

TABLE 4 (CONTINUED)

OXYGEN USAGE-EPA-DC PILOT PLANT
1971

	Primary Eff. COD (lb/mil.gal)	Final Eff. COD (lb/mil.gal)	Waste Sludge COD (lb/mil.gal)	COD Removal from System (lb/mil.gal)	Nitrate Nit. Demand (lb/mil.gal)	Exhaust Oxygen (lb/mil.gal)	Final Eff. D.O. (lb/mil.gal)	Total	Oxygen Suspended (lb/mil.gal)
March 1-18	2050	630	160	1260	0	160	30	1450	1450
April	2180	380	700	1100	0	130	40	1270	1300
May	2500	450	890	1160	0	40	50	1250	1260
June	1750	340	80	1330	0	80	60	1470	1600
July	1940	270	270	1400	60	300	60	1820	2200
August	1830	260	270	1300	70	260	40	1670	1690
September	1920	280	400	1240	270	200*	40	1750	±
October	1880	310	390	1180	200	200*	30	1610	±
November	1950	440	220	1290	220	200*	40	1750	1740
December 1-21	1990	490	230	1270	190	200*	40	1700	2000
December 22-31	1970	440	300	1230	70	150*	40	1490	1450

TABLE 4 (CONTINUED)

OXYGEN USAGE-EPA-PILOT PLANT

1972

	Primary Eff. COD (lb/mil.gal)	Final Eff. COD (lb/mil.gal)	Waste Sludge COD (lb/mil.gal)	COD Removal from System (lb/mil.gal)	Nitrate Nit. Demand (lb/mil.gal)	Exhaust Oxygen (lb/mil.gal)	Final Eff. D.O. (lb/mil.gal)	Total	Oxygen Suspended (lb/mil.gal)
January 1-19	2160	390	630	1140	40	40	40	1260	1410
February	1870	450	270	1150	0	63	50	1263	±
March 1-17	1990	675	340	975	24	37	50	1076	1120
March 21-31	1920	465	380	1075	25	25	50	1175	1175
April	1970	400	440	1120	49	37	37	1243	1250
May	1900	400	250	1250	266	40	14	1570	1540
June	1840	360	50	1430	360	170	36	1996	1850

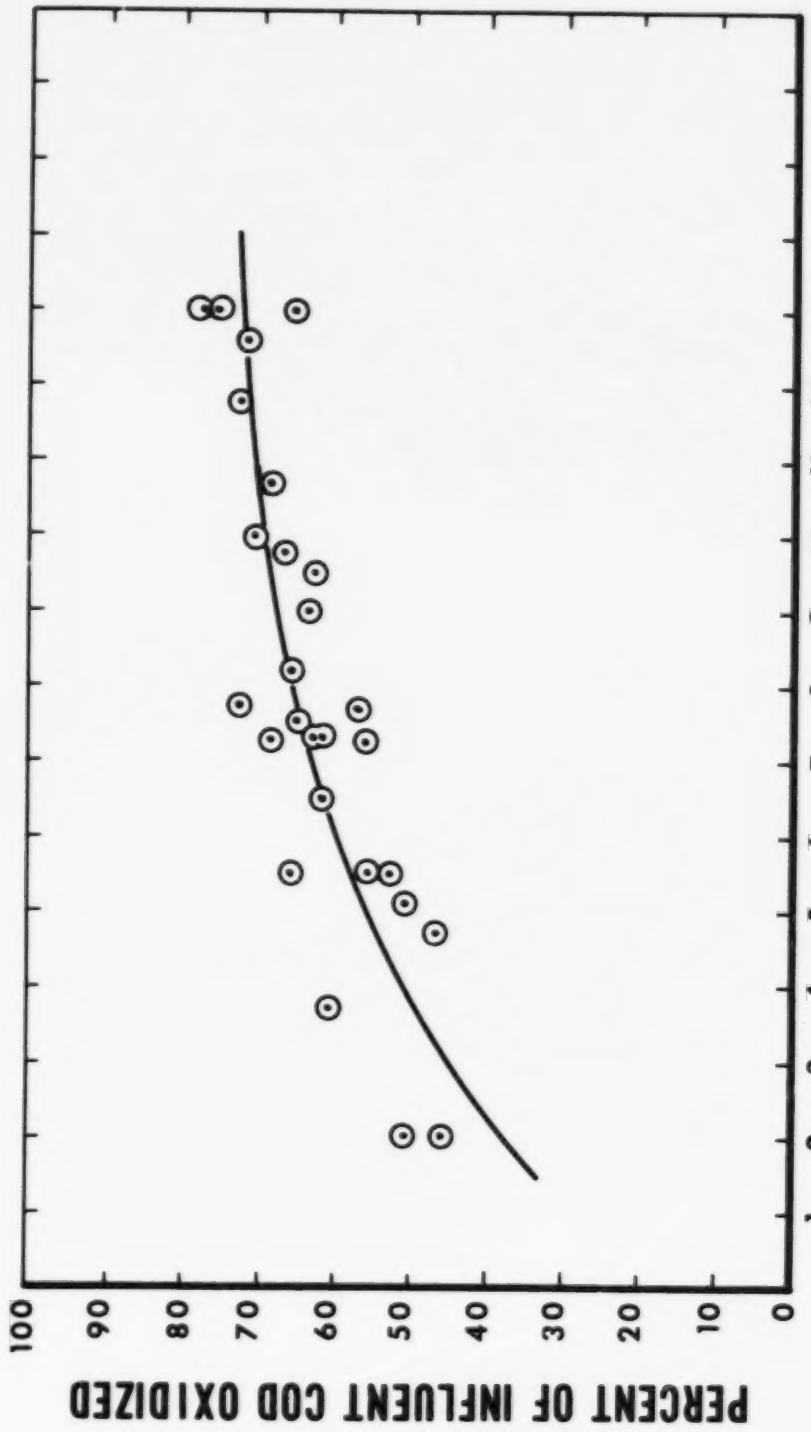

Figure 5. Percent COD Oxidized as a Function of SRT

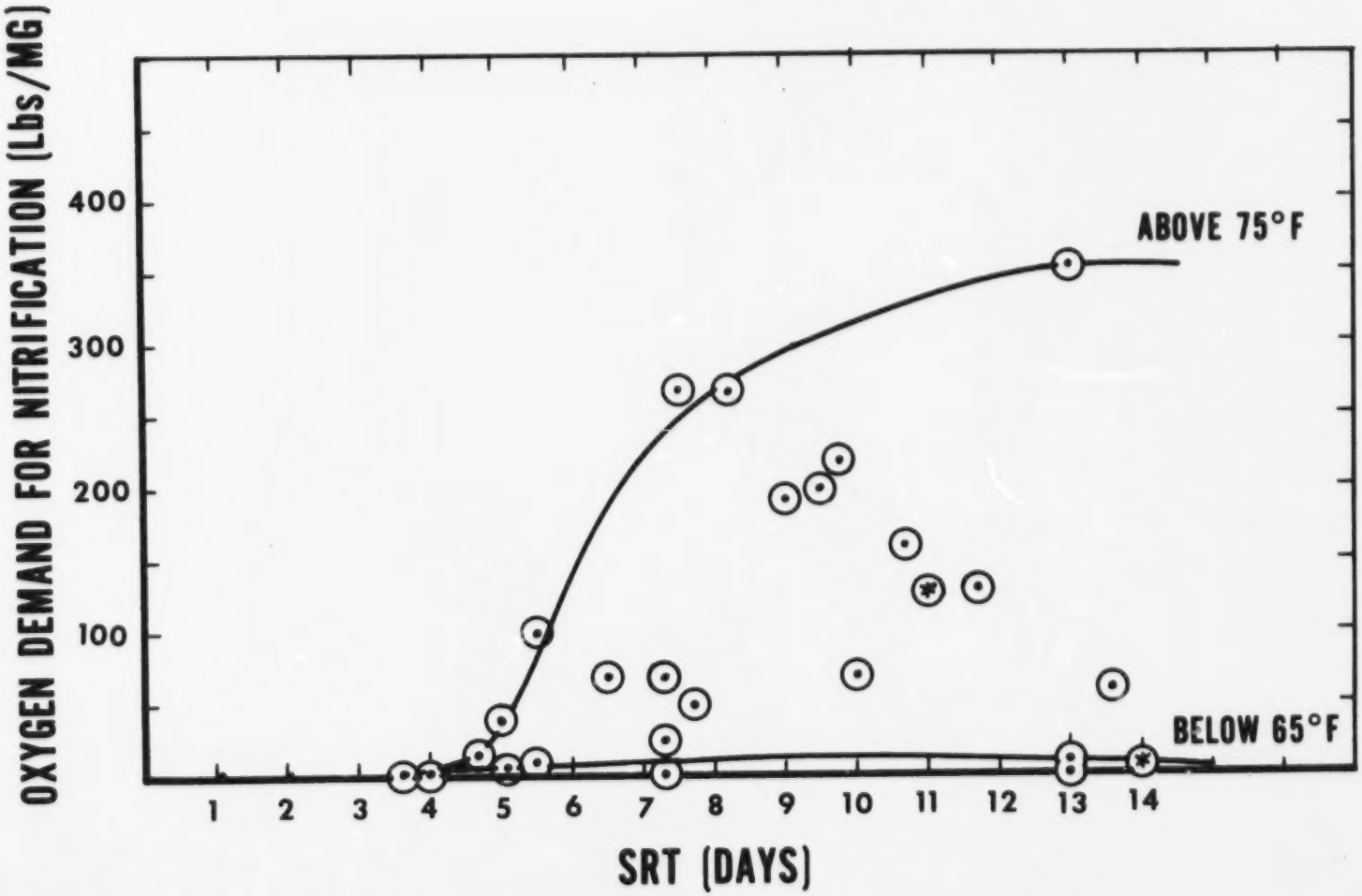

Figure 6. Nitrification Oxygen Demands

SECTION VI
CLARIFICATION OF OXYGEN ACTIVATED SLUDGE

The second important aspect in the oxygen aeration system is liquid/solids separation. At the EPA-DC Pilot Plant, as at Batavia, gravity clarification is employed. As mentioned before, soluble residual BOD in the effluent averaged less than 5 mg/l in the test periods indicating virtually complete BOD insolubilization. Thus, most of the residual BOD in the Blue Plains oxygen system effluent was associated with suspended solids and overall removal of suspended solids and BOD were a function of clarification efficiency.

Clarifier efficiency is in turn a function of the basic settling characteristics of the solids as well as of the actual design and operation of the clarifier. With the normally higher mixed liquor concentrations used in the oxygen aeration process, design criteria of both clarifiers (i.e., overflow rates and volume) and thickeners (i.e. solids loading - $lb/ft^2/day$) should be considered. The Ten State Standards (15) suggest that conventional activated sludge clarifiers be designed for average overflow rates of 800 gpd/ft^2. The Water Pollution Control Federation Manual of Practice (1959) (16) suggests that the solids loading be held below a peak of 30 $lb/ft^2/day$. Overflow rates and solids loading criteria must be better defined for high solids systems such as oxygen aeration.

The settling of a particle in a fluid (such as mixed liquor solids in water) is dependent on gravitational force. The magnitude of the driving force for settling depends on the difference in the densities of the bulk particle and the fluid. The density of the bulk mixed liquor particle is determined by the composite of bacterial mass, bound water and miscellaneous flocculated debris or chemical precipitant. Furthermore, by definition, the driving force is affected by changes in the specific gravity of water (caused either by variations of temperature of dissolved solids concentration).

Viscosity shear force is defined as the retardant of drag force in settling. The magnitude of the drag force is dependent on the viscosity of the fluid (in this case water) the shape of the particle and the velocity of the particle relative to the fluid. Viscosity of water is likewise dependent on temperature. The particle shape affects the streamlines (hydraulic pattern) around the particle. Consequently, a well rounded mixed liquor particle will settle faster than an irregular shaped particle, such as those found in a filamentous culture. As mentioned previously, the velocity of the particle relative to the fluid also affects the drag force. The drag force increases as the velocity of the fluid relative to the particles increases. Consequently as concentration of the bulk particles increases, the net particle velocity

is reduced due to the increased velocity of the fluid as it is forced to pass through an increasingly smaller crossectional area.

In summary, the basic settling characteristics of the mixed liquor typically are a function of:

1. Concentration of the mixed liquor
2. Particle shape
3. Particle density
4. Seasonal variation
   a. Physical changes in water density and viscosity with temperature
   b. Metabolic changes with temperature
   c. Seasonal loading variation

Duncan and Kawata (17) and Dick (18) observed that in the range of initial solids concentration when subsidence settling occurred, a straight line relationship exists between the log of the initial settling velocity and the log of the initial concentration. Consequently an equation in the form of:

$$Vi = a\ Ci^{-n}$$

where Vi = initial settling velocity
Ci = initial concentration
a = constant of magnitude
n = constant of rate of change

could be established for each sludge. Furthermore, a relationship of this type can be determined for each of the three types of initial settling conditions (free particle settling, subsidence and consolidation settling).

As seen in Figure 7, the log of the initial settling rate is a function of the log of the solids concentration. This sample curve illustrates that two relationships exist. The first at lower MLSS levels corresponds to free particle settling and is characterized by the absence of an initial discrete subsiding interface and a zone of homogeneous settling solids. The second at higher MLSS levels has both an initial discrete interface and a zone of homogeneous settling particles (zone settling). Thus, the sizing of a clarifier is a function of the MLSS concentration and must be coordinated with the reactor (and the sludge handling facilities) to achieve the desired biological capabilities of the system.

Another important factor is the particle shape. Normally, as shown in Figure 2, the oxygen mixed liquor particles have rounded shapes. However, if filamentous growth exists, as experienced below an SRT of 5 days in D.C. (air and oxygen), both settling rates and compaction deteriorate. The range, if any, that filamentous growth appears is

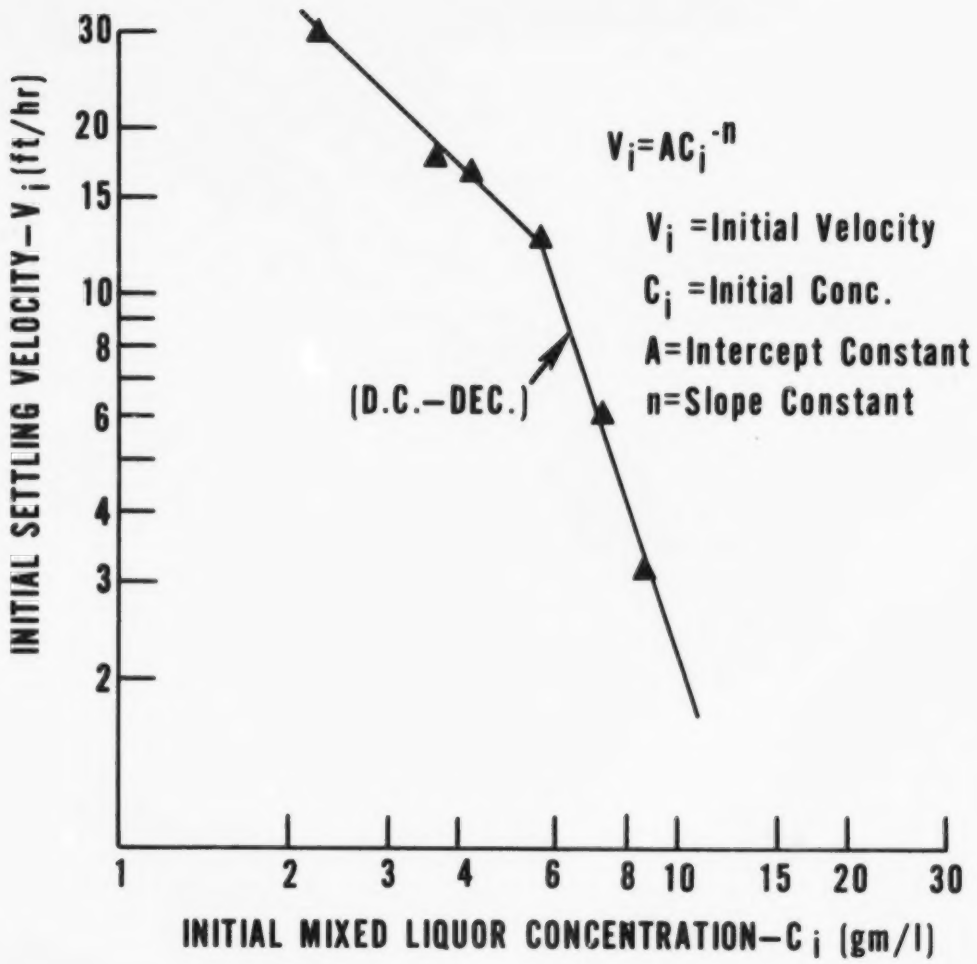

Figure 7. Initial Sludge Settling Velocity as a Function of Concentration

unique to each location and should be defined for that location. The presence of industrial fibers also had an effect similar to filamentous growth on mixed liquor settling characteristics.

Still another important factor in the basic settling characteristics is the density of the particles in relationship to that of the water. It is the difference in density that is the driving force for settling. The VSS/TSS ratio (or volatile %) is one relative indication of density. There are several ways to improve the particle density. One is to feed raw wastewater instead of primary effluent to the oxygen aeration system, thus incorporating the normally denser particles captured in primary sedimentation into the biomass, such as occurred at Batavia (7). Again, the sizing of the reactor oxygen supply, etc. must be compatible with the increased organic loading. In Washington, heavy rains and unusually high flows wash silt and clay into the sewer system. These materials subsequently become incorporated in the mixed liquor solids and have increased sludge settling rates 30% to 60% (Figure 8).

In like fashion, operating under different biological conditions can alter the sludge settling characteristics and the sludge volume index (SVI). At SRT maintained below 5 days in Washington, D.C., filamentous bulking occurs in oxygen and air systems and increases the SVI. In addition, as the system is operated at SRT's above 10 days, bio-flocculation decreases and some affects on settling and SVI are observed (Figure 9). Also, in temperatures below 70°F at SRT's around 7 days, the SVI's increased. Grab samples of MLSS analyzed for the hexane soluble grease (13) reveal high levels around 6-9% by weight during these periods. Conclusive evidence needs to be obtained on the relation between settling rates, grease concentration, and temperature (i.e. determine if and when grease accumulations occur and the effect on the SVI and settling rate).

The mode of operation of the clarifier affects the density and settling characteristics of the oxygen sludge. Two major methods of clarifier operation are available. One is to use the blanket as a filter and the other is to permit classification of the settling solids. The first method can be accomplished in two different ways: (1) by providing sufficient depth to the clarifier such that the mixed liquor passes up through the clarifier blanket (which acts as a filter); (2) by carrying high MLSS concentrations (usually above 4000 mg/l) such that the particles settle in a subsidence (zone) settling pattern with discrete interfaces existing between the homogeneous subsiding particles and the decant. In the subsidence zone, the relatively uniform concentration of particles are nearly homogeneously mixed by the countercurrent turbulence produced by water passing around the solids. The homogeneous subsiding blanket does not allow classification of individual particles because the settling blanket acts as a filter.

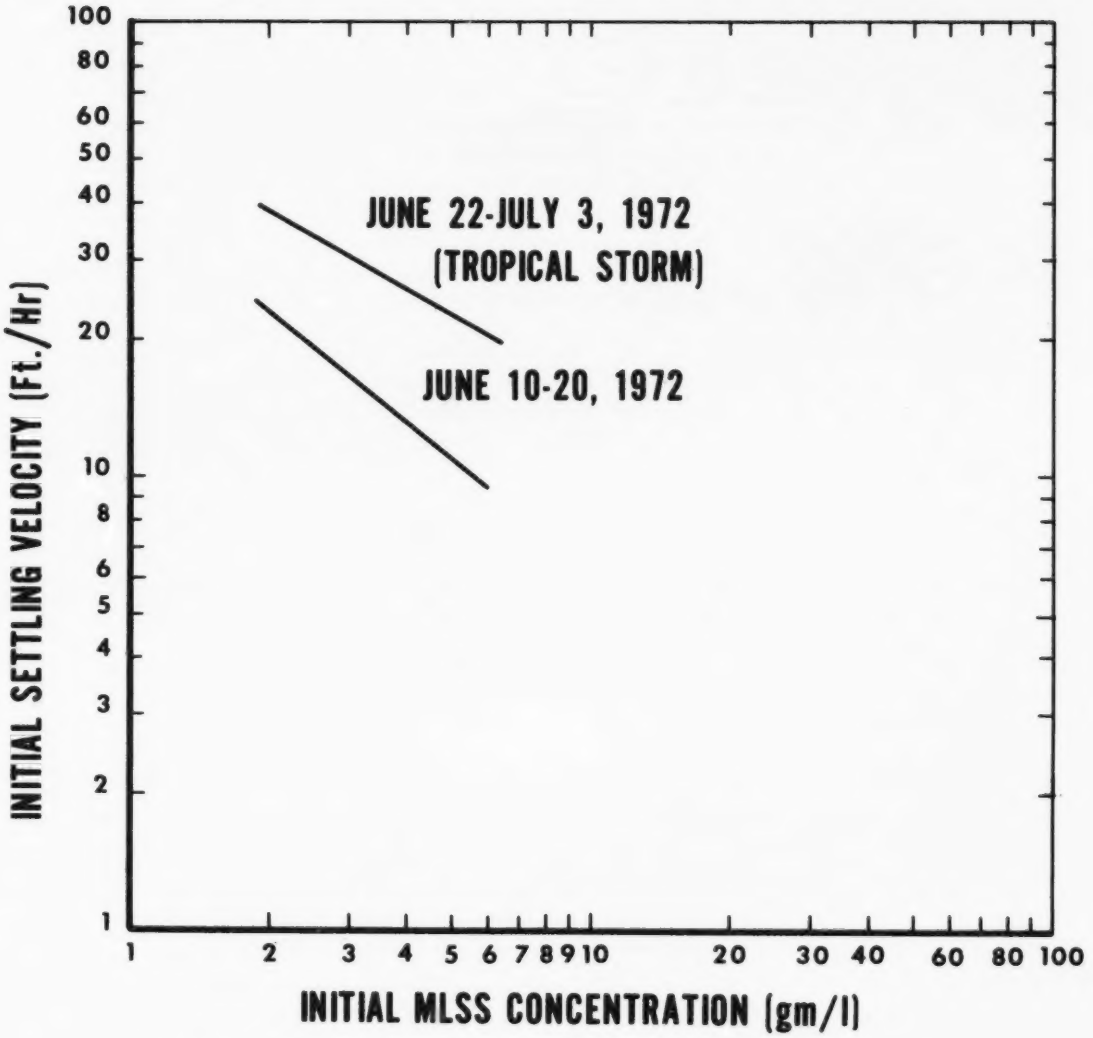

Figure 8. Effect of Rains on the Initial Sludge Settling Velocity

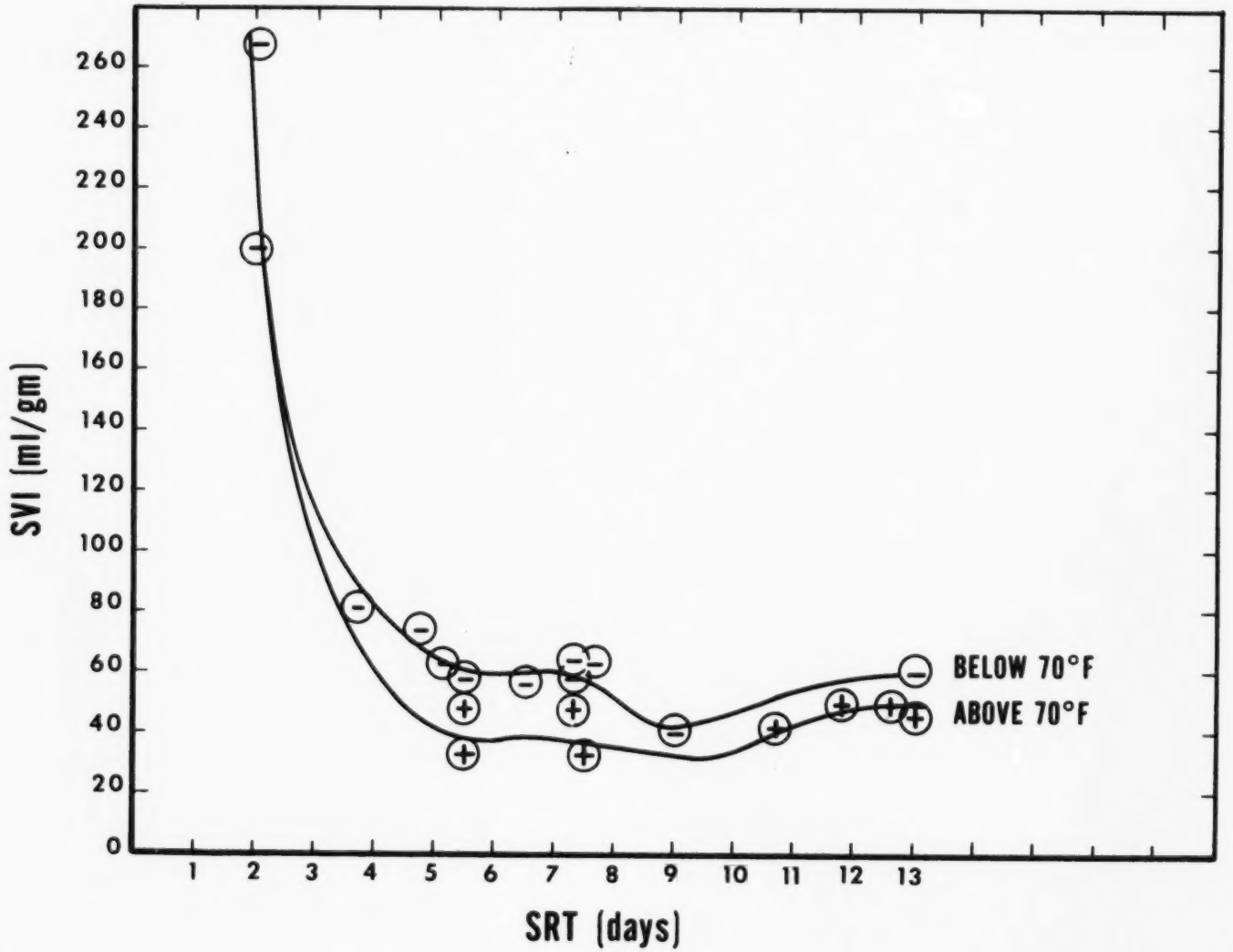

Figure 9. Effect of SRT on SVI

At Washington, with MLSS concentrations below 4500 mg/l, subsidence (zone) settling does not occur during the initial portion of settling. This provides for a second method of clarifier operation where classification of the discrete settling particles can occur if the mixed liquor is fed above the clarifier blanket level. The lighter or unsettlable particles, thus, can be purged from the system. The effluent suspended solids accordingly increased from 15 mg/l to 25 mg/l during this method of operation with a corresponding increase in effluent BOD.

Seasonal variations also effect sludge settling characteristics in oxygen as well as in air systems. These variations become critical as the MLSS of the mixed liquor increases. The pure physical changes in the wastewater density and viscosity contribute to slower settling rates as the wastewater temperatures decreases. As the density of the water increases, the driving force for settling (which is the difference in density between water and the settling particles) decreases for a similar particle density. The drag force, viscosity, also increases with decreasing temperature and contributes to slower settling rates in colder waters. Figure 10 shows a series of liter batch settling tests conducted in June, 1971 in which the temperature of the mixed liquor was lowered. As expected, the colder samples settled slower. In Figure 11, the batch flux (concentration multiplied by settling velocity) or the solids loading in $lb/ft^2/day$ is shown for the previous tests. Solids loading is often used in thickener design. Again, the effect of wastewater temperature is evident.

Besides the physical changes caused by seasonal variation, another factor which must be considered is the metabolic change brought about by changing wastewater temperature. Figure 12 shows that the settling characteristics of oxygen mixed liquor change seasonably at D.C. At similar SRT's, the initial settling rate in a 1 liter graduated cylinder test decreased from approximately 10 ft/hr to 7 ft/hr at a concentration of 6000 mg/l as the temperature changed from 81°F to 71°F. The clarifier was being operated to capture unsettleable particles at this time. In Figure 13, the clarifier was operated to purge the unsettleable particles; but, again the solids showed a decreasing initial settling rate with decreasing wastewater temperature for a similar biology. The initial settling rate in the 1 liter test decreased from 70°F to 63°F. Likewise, as the wastewater temperature increased, the settling rates increased in the Spring and Summer (Figure 14).

The D.C. settling characteristics of oxygen activated sludge conform to the range of settling rates published by Wilcox (19) (Figure 15) when non-filamentous cultures (less than 200 SVI) are tested on a 6" stirred column. With filamentous growth the settling rate markedly

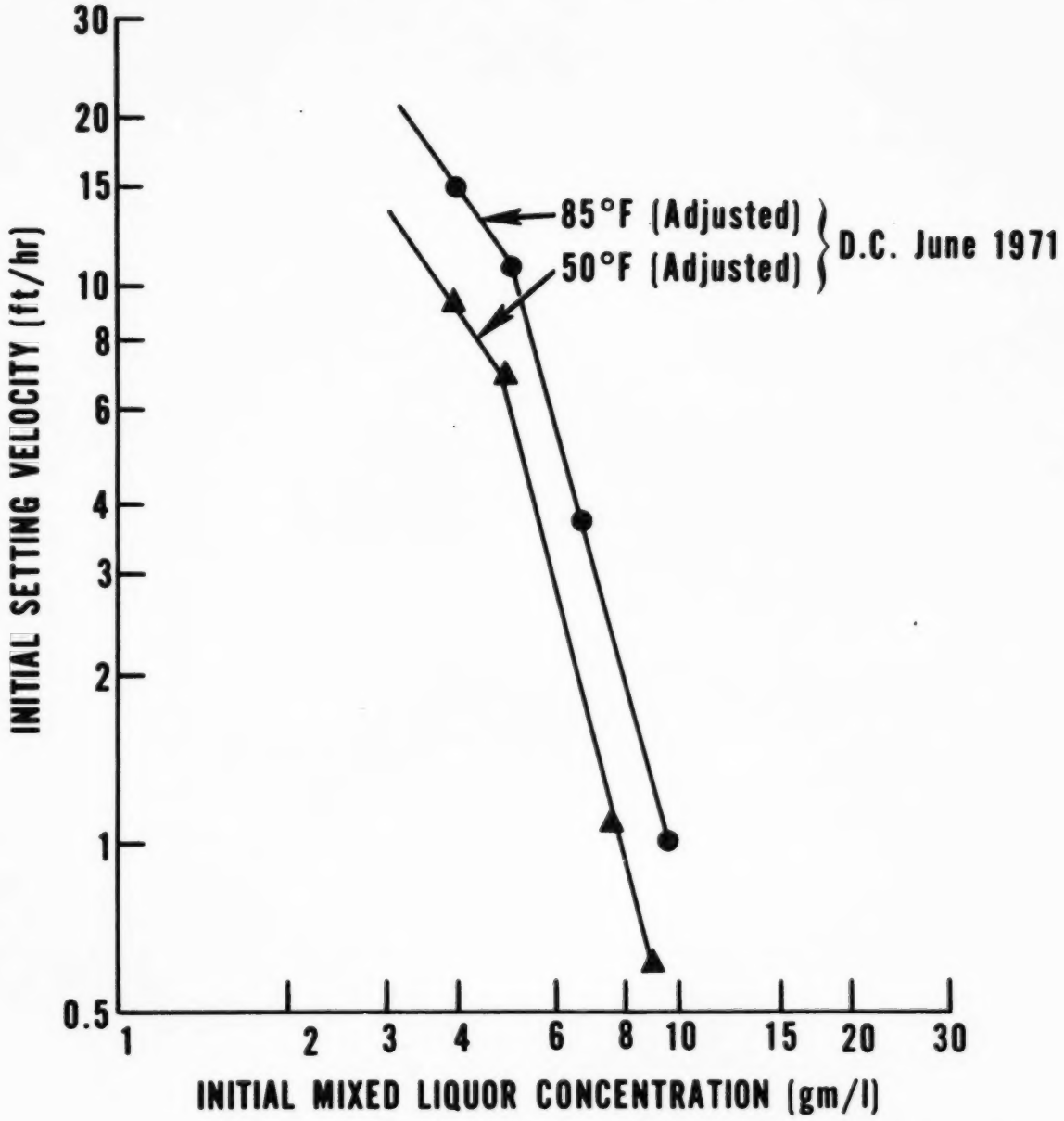

Figure 10. Effect of Temperature Adjustments on Initial Sludge Settling Rates

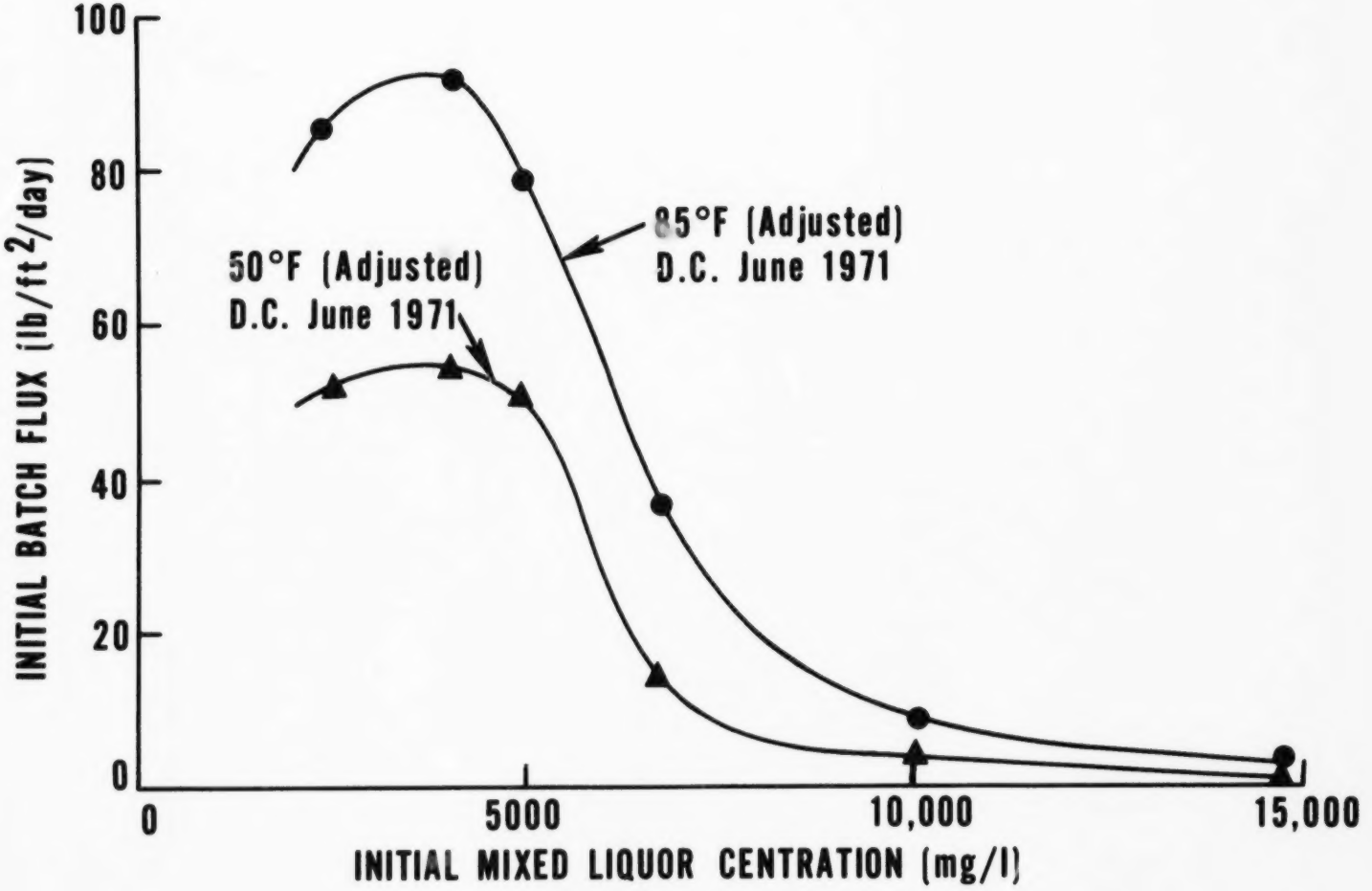

Figure 11. Effect of Temperature Adjustments on Initial Batch Flux

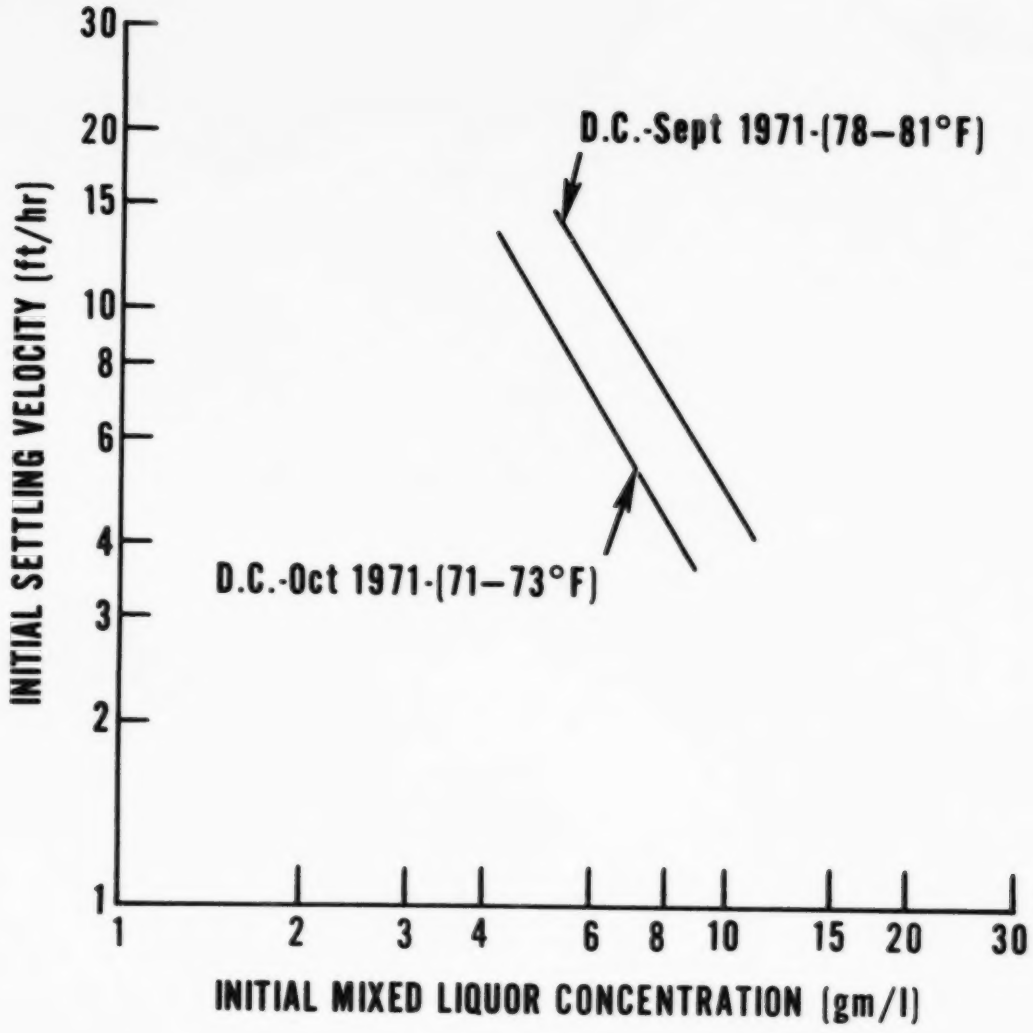

Figure 12. Effect of Wastewater Temperature on Oxygenated Sludge Settling Rates (Sept.-Oct. 1971)

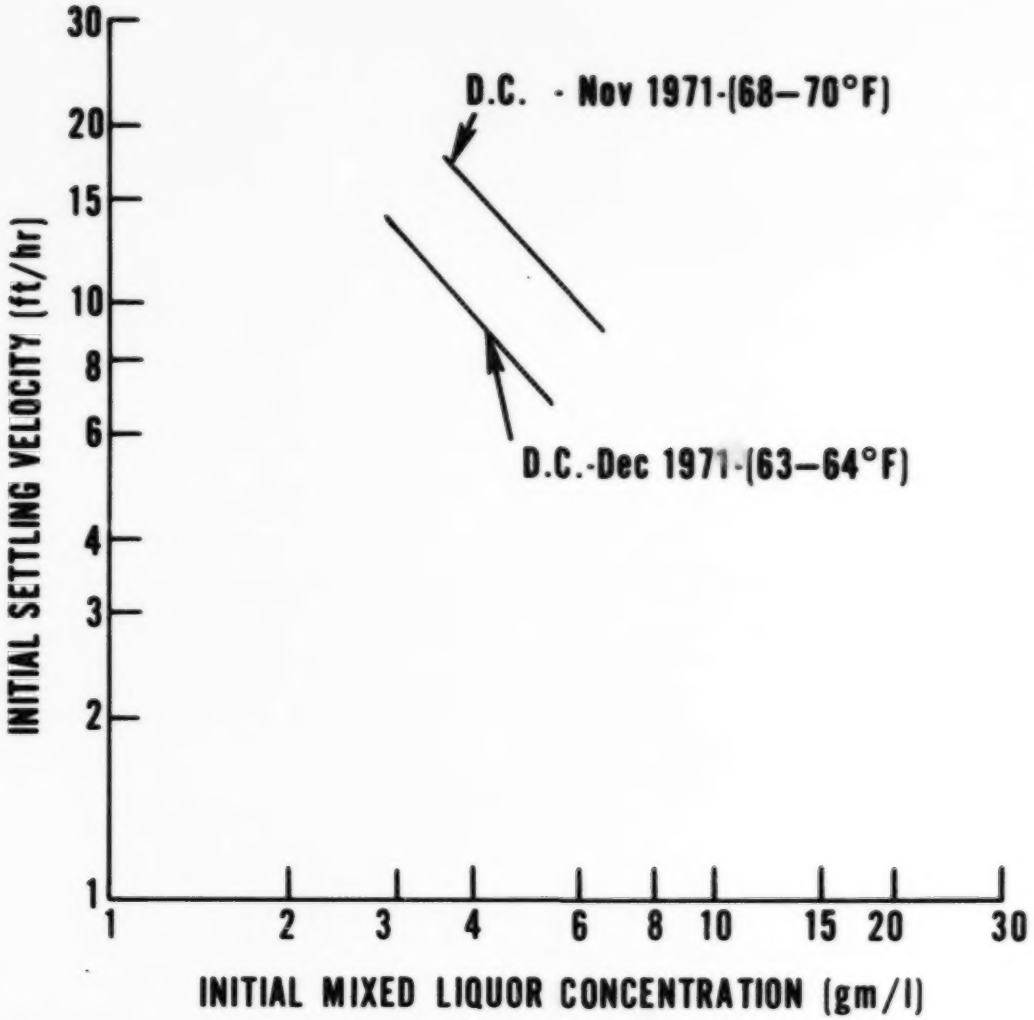

Figure 13. Effect of Wastewater Temperature on Oxygenated Sludge Settling Rates (Nov.-Dec. 1971)

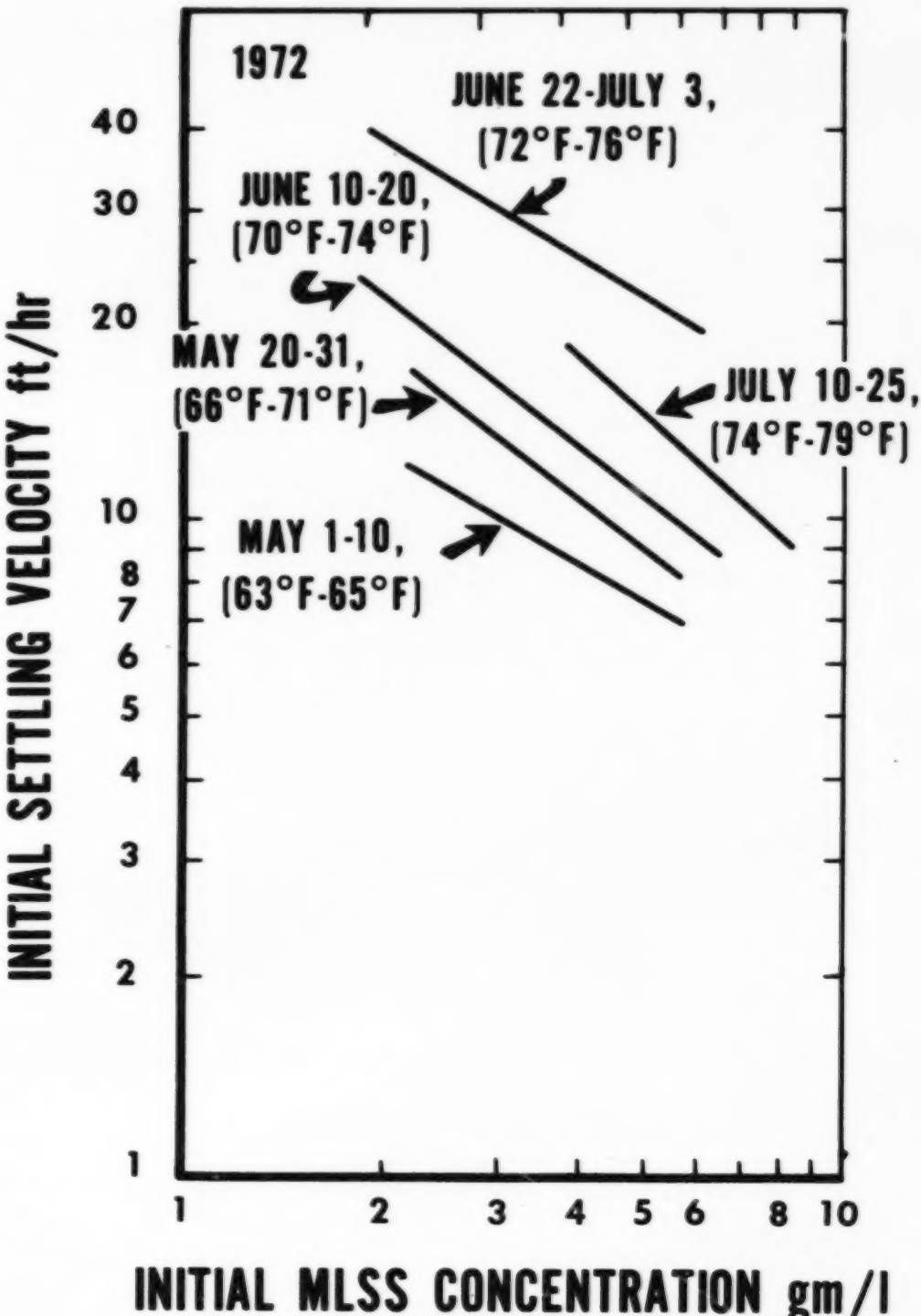

Figure 14. Effect of Wastewater Temperature on Oxygenated Sludge Settling Rates (May-July 1972)

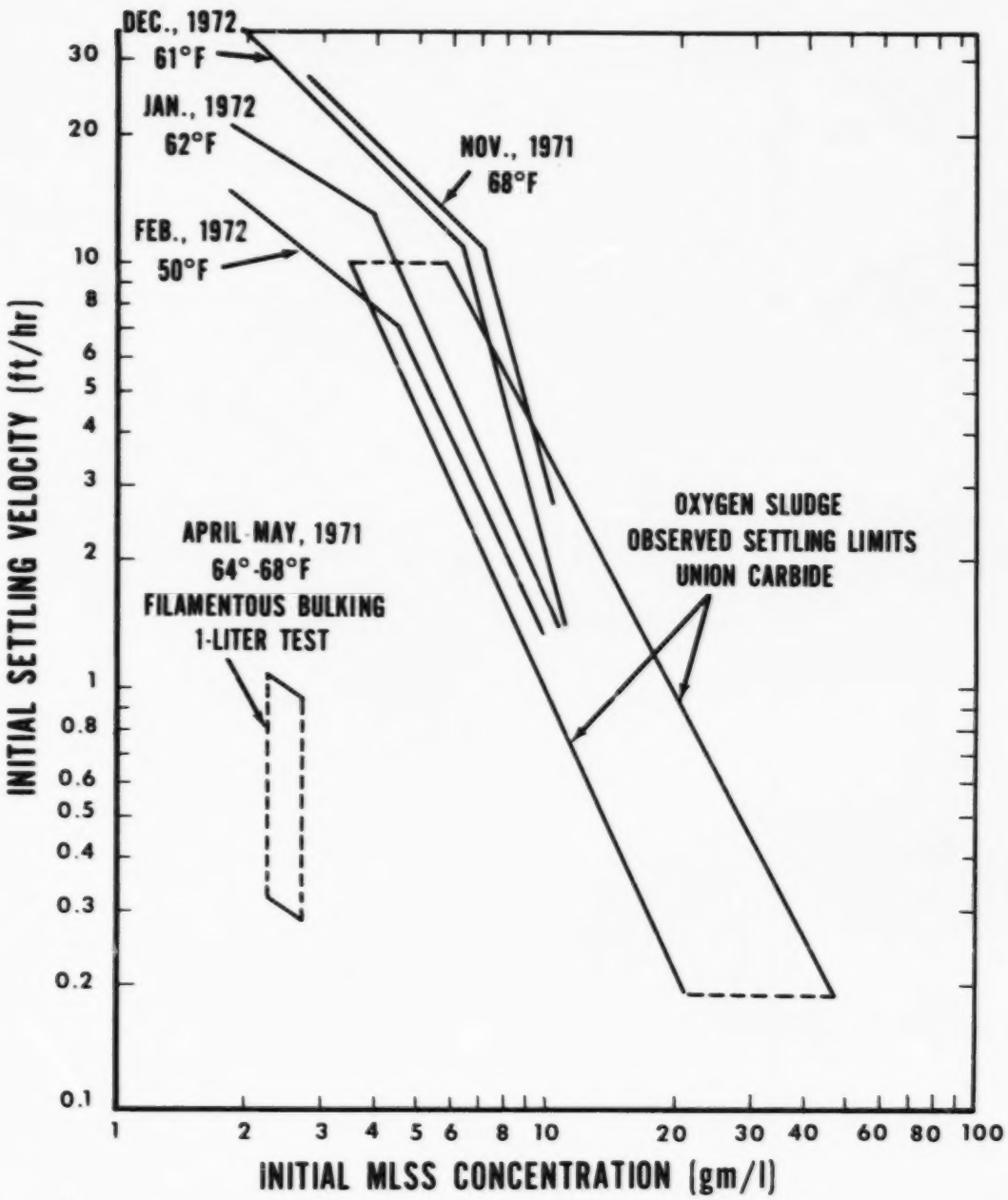

Figure 15. Initial Settling Velocities of Oxygen Sludges as a Function of Concentration

deteriorated to as little as 0.3 to 1.0 ft/hr at MLSS of 2500 mg/l in the standard 1 liter settling test. Unfortunately, 6" stirred settling data (minimized wall affects and particle bridging) was not obtained.

The air activated and oxygen activated sludge settling rates are comparable. As seen in Figure 16, the air activated sludge settling rates observed by Agnew (2), Reed and Murphy (21), Dick (22), Javaheri (23), Eckenfelder (24), Rudolfs (25), and Flischerstorm (26) fell in the same settling range as Wilcox and EPA-DC oxygen sludges for non-bulking sludges. Further, the settling rates observed in the nitrification and denitrification sludges of a parallel EPA-DC multi-stage mineral addition pilot plant fell in the same ranges as the air and oxygen sludges (Figure 17). Similar patterns of decreasing settling rates with decreasing wastewater temperature were also observed in nitrifying and denitrifying mixed liquors.

Clarifier operation and design are equally important to the basic settling characteristics of the solids in gravity clarification. Besides selecting an overflow rate compatible with reactor sizing, the depth and method of clarifier feed are important as discussed earlier for high solids capture or for solids classification. Other important design considerations are the volume or detention time of the clarifier and recycle rate. At Blue Plains, the oxygen system underflow solids concentration varies between 1.0% and 1.4% with an average clarifier detention time of 1.9 hours. With 2.8 hours average detention time, the underflow solids concentration rises to 2.0% - 2.4% with similar recycle rates. For any selected F/M, the clarifier volume will thus affect the under flow concentration and thus the sludge recycle rate.

Inventory solids are another consideration in clarifier operation. The total solids inventory is a result of both the build-up of solids in the blanket and the solids actually in the transition (or settling) process. A simple increase in the wasting rate reduces the blanket level if the increase is caused by the build-up of solids. However, as the initial settling rates decreases, as in Washington, for a given MLSS concentration, the sludge inventory also increases in the clarifier as more solids are in transition (or settling). In this case, increased wasting rates does not reduce the solids inventory in the clarifier without thinning the MLSS or altering the biology. It appears that the increased inventory can be a result of slower settling solids as well as a backlog of solids due to inadequate wasting.

In the summer of 1970 at an MLSS concentration in excess of 8,000 mg/l, peak clarifier overflow rates of 1940 gpd/ft^2 were observed on the Blue Plains oxygen system as shown in Table 5. During the 1970-71 winter, the peak sustained overflow rates which could be maintained without the blanket coming over the weirs was 975 gpd/ft^2 at MLSS concentrations

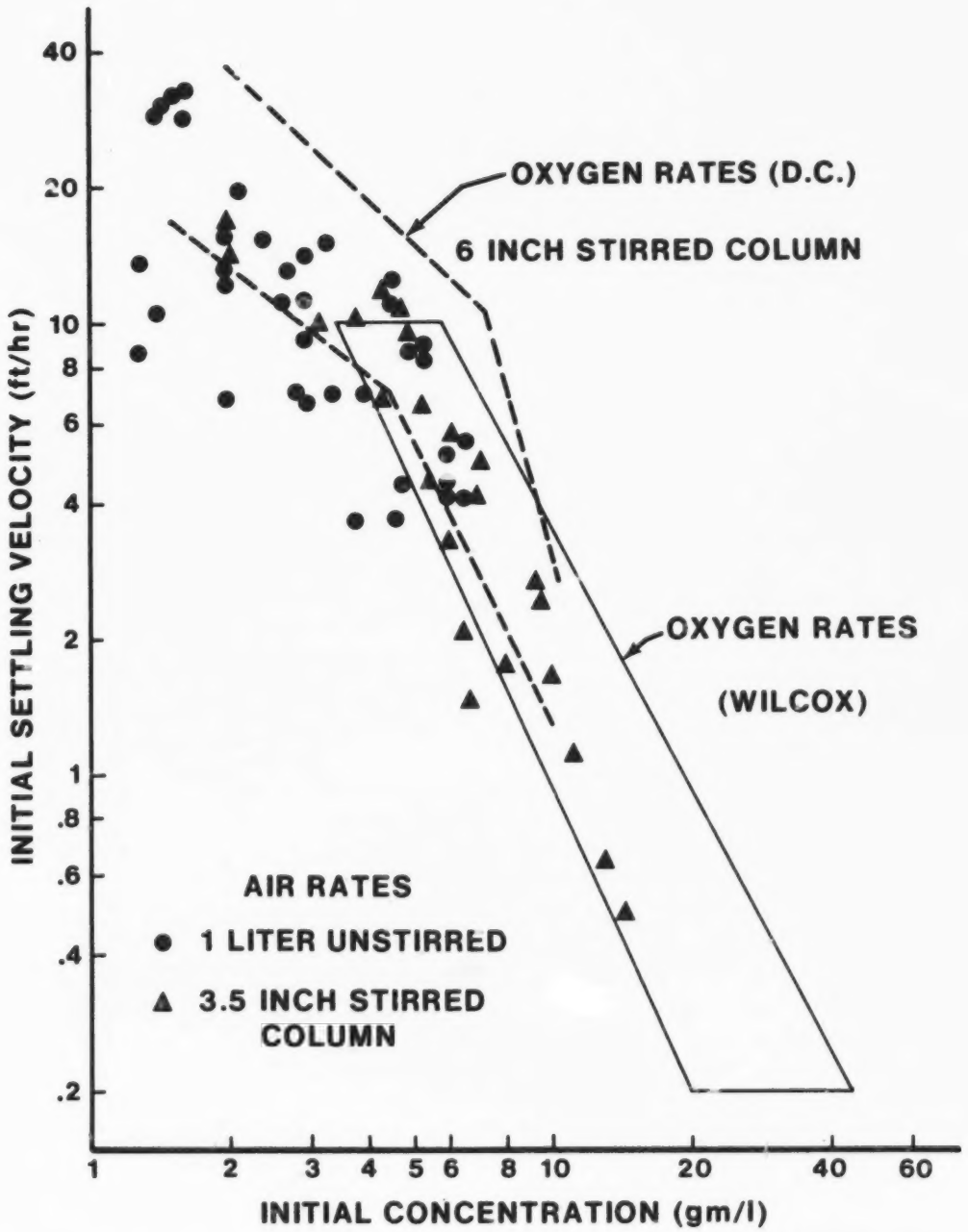

Figure 16. Initial Settling Velocities of Oxygen Sludges Compared with Air Sludges

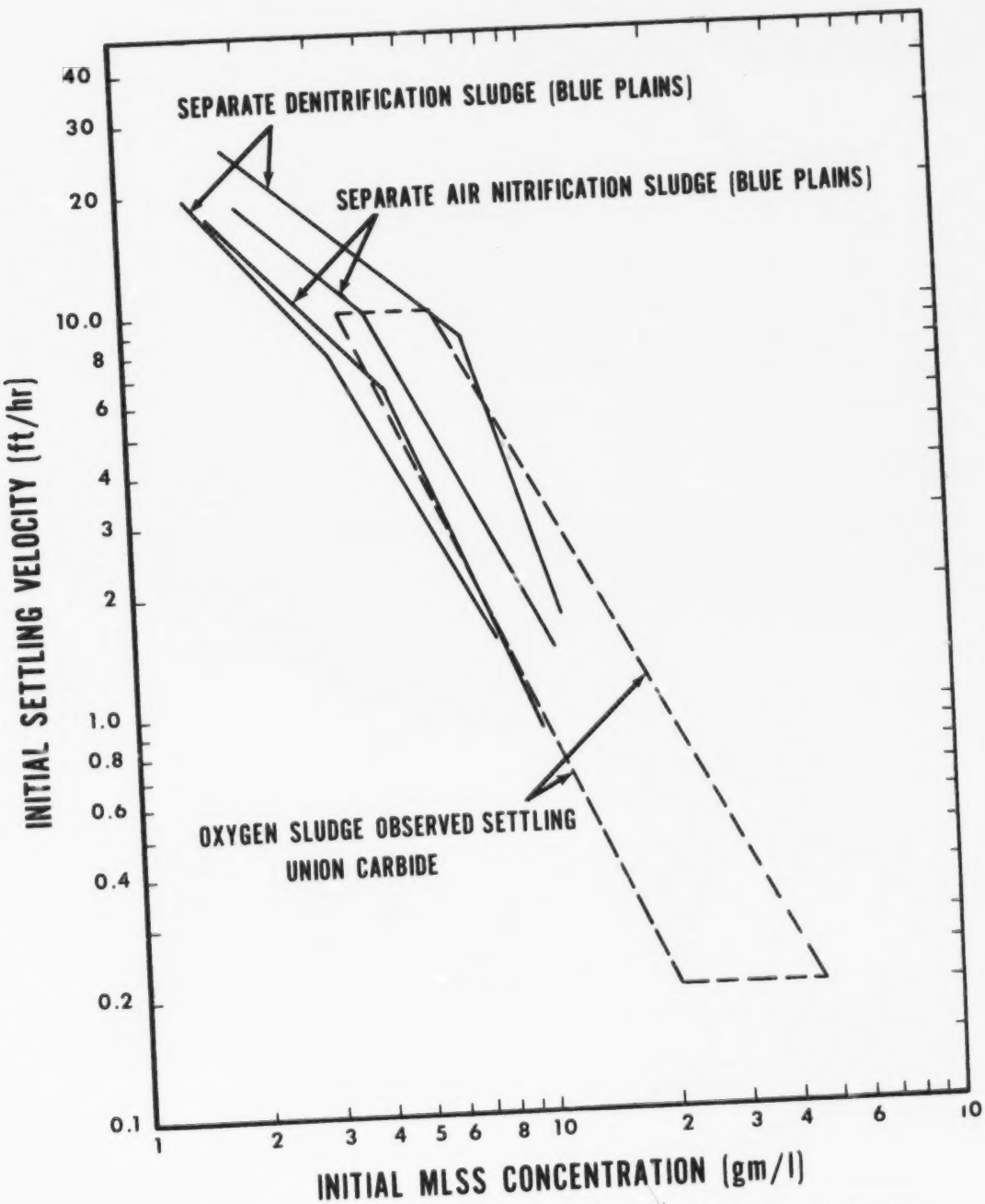

Figure 17. Observed Initial Settling Velocities of Oxygen Sludges Compared with Nitrification and Denitrification Sludges

TABLE 5
CLARIFIER VARIABLES AND OPERATING CONDITIONS
EPA-DC PILOT PLANT
1970

	Ave. Overflow Rate (gpd/ft^2) At Surface	Peak Overflow Rate (gpd/ft^2) At Surface	Ave. Solids Loading (lb/day/ft^2)	SVI	Underflow Solids % Dry Solids	Underflow Solids % Volatile	Waste Solids lb/mil gal	Volatile lb/mil gal	Effluent Solids lb/mil gal	Effluent volatile lb/mil gal
June 12-30	750 +	750 +	37	80	1.16	75	161	121	296	198
July	1210 +	1210 +	75	48	1.34	70	40	28	445	245
August 1-25	1210 +	1210 +	58	50	1.27	75	144	108	166	113
September	1280 ±	1940	61	42	1.40	80	250	200	204	141
October 3-21	1280±	1940	88	33	2.14	65	680	441	290	189
November 10-30	1280 ±	1940	68	48	1.40	81	230	202	470	342
January 1-16	975 ±	975	55	60	1.08	90	193	174	197	118
January 17-31	975 ±	975	42	73	1.00	80	253	202	483	400

TABLE 5
CLARIFIER VARIABLES AND OPERATING CONDITIONS
EPA-DC PILOT PLANT
1971

	Ave. Overflow Rate (gpd/ft^2) At Surface	Peak Overflow Rate (gpd/ft^2) At Surface	Ave. Solids Loading (lb/day/ft^2)	SVI	Underflow solids % Dry Solids	Underflow solids % Volatile	Waste Solids lb/mil gal	Volatile lb/mil gal	Effluent Solids lb/mil gal	Effluent volatile lb/mil gal
March 1-18	950 ±	950	35	81	0.79	77	130	100	410	375
April	(290-620)	(290-620)	17	120-190	0.85	80	-	-	-	-
May	560	560	21	265	0.78	77	720	550	100	-
June	(280-650)	(280-650)'	24	173	1.28	73	142	104	88	-
July	650	975	54	50	1.92	70	168	118	92	-
August	650	975	61	30-35	2.22	70	253	178	133	100
September	650	975	58	33	2.38	70	460	323	160	104
October	650	975	47	42	2.26	71	340	242	130	99
November	650	975	36	36	2.05	79	178	140	194	130
December 1-21	650	975	33	40	2.41	81	200	160	190	123
December 22-31	650	975	32	56	2.30	82	310	254	140	88

TABLE 5
CLARIFIER VARIABLES AND OPERATING CONDITIONS
EPA-DC PILOT PLANT
1972

	Ave. Overflow Rate (gpd/ft^2) At Surface	Peak Overflow Rate (gpd/ft^2) At Surface	Ave. Solids Loading (lb/day/ft^2)	SVI	Underflow solids % Dry Solids	Underflow solids % Volatile	Waste solids lb/mil gal	Volatile lb/mil gal	Effluent Solids lb/mil gal	Effluent volatile lb/mil gal
January 1-19	170-740	170-740	9-40	58	2.15	85	530	450	100	80
February	520	760	30	58	2.00	79	180	143	290	194
March 1-19	520	760	29	62	1.63	75	310	230	435	330
March 21-31	520	760	33	64	1.49	76	335	255	105	67
April	520	760	32	63	1.37	70	440	310	185	140
May	480	700	27	62	1.17	75	222	166	189	148
June	520	760		46	1.40	72	45	31	165	136

that varied from 3900 to 5300 mg/l. The causative agents which reduced the allowable overflow rates for satisfactory operation from summer to winter were undoubtedly a combination of all the above mentioned factors, especially the decreased wastewater temperature. As expected with filamentous growth in late spring of 1971, allowable clarifier overflow rates were markedly decreased as shown in Table 5. Again, during the current 1971-72 winter, maximum overflow rates of 975 gpd/ft^2 were demonstrated at Washington.

Air floatation was tried as an alternate method of clarification. Initially the flow from the reactor was split between an air floatation unit and a gravity clarifier. The solids thickened to 5-6% in air floatation but averaged 3.82% in the recycle as additional water was brought over with skimming (Table 6). The parallel gravity clarifier recycle averaged 2.31%. The effluent suspended solids from air floatation averaged 34 to 41 mg/l at a solids loading of 14-15 lb/ft^2/day and the effluent suspended solids from gravity clarification averaged 7 to 21 mg/l at a solids loading of 36-37 lb/ft^2/day. Later the loadings on the air floatation unit were increased to an average of 25 lb/ft^2/day (400 gpd/ft^2, overflow rate) with peaks of 36 lb/ft^2/day (580 gpd/ft^2 overflow rate) and the suspended solids further deteriorated to an average of 74 mg/l even with 0.75 mg/l of Dow C-31 polymer.

In general air floatation was not a suitable alternative to gravity clarification because the effluent suspended solid were much higher than a comparable gravity settler.

TABLE 6

AIR FLOTATION CLARIFICATION

	Overflow Rate, gpd/ft^2	Solids Loading $\#/ft^2/day$	Effluent S.S. mg/l	Recycle %
July 24-31, 1972				
Gravity Clarifier	520	37	7	2.31
Air Flotation	200	15	34	3.82*
August 1-20, 1972				
Gravity Clarifier	520	36	21	2.40
Air Floatation	200	14	41	3.26*
August 21-31, 1972				
Air Floatation	400	25	74	2.10*

*Grab samples before skimming were 5-6%, however, increasing skimming to improve effluent suspended solids diluted the floated solids concentration.

SECTION VII
SOLIDS PRODUCTION AND HANDLING OF WASTE OXYGEN ACTIVATED SLUDGE

The other integral part of the oxygen system is the excess solids handling equipment. The important factor in the solids handling equipment is the relative ease with which the oxygen activated sludge process can be operated in endogenous respiration, thereby substantially reducing the quantity of excess sludge to be handled. This capability permits the reduction in the number and/or size of the selected sludge handling and disposal facilities. However, the increased operating costs resulting from the increased oxygen necessary to bio-oxidize (burn-up) the excess sludge and the larger reactor/clarifier capacities needed to hold the increased solids inventory required for endogenous respiration must be balanced economically with the reduction in size of the solids handling and disposal units.

The first step in solids handling is thickening. The oxygen activated sludge was thickened to 4-5% by either gravity or air floatation without chemical additives. The Komline Sanderson HR air floatation thickener (1 ft^2) yielded greater than 4% solids at loadings up to 6 $lb/ft^2/hr$ (Figure 18) with effluent suspended solids concentrations less than 200 mg/1. When the oxygen activated sludge solids were mixed with primary solids in weight ratios of 0.2 secondary to 1.0 primary and 0.4 secondary to 1.0 primary, the sludge concentration gravity thickened to as high as 11% solids.

Dewatering studies were performed by vacuum filtration, centrifugation, and filter pressing. The oxygen waste sludge dewatered in vacuum filter tests to 8-12% with polymer conditioning and to 13 - 21% with chemical conditioning (ferric $FeCl_3$ at 5% and lime at 16-20% based on dry weight). Vacuum filter leaf tests showed that the 35% solids, obtained at form rates of 21.0 $lb/ft^2/hr$ for primary solids alone, decreased with increasing activated sludge content to 20.7% solids at form rates of 11.0 $lb/ft^2/hr$ for a 1:1 ratio of primary to oxygen sludge (Figure 19 and 20). In the 3 ft by 1 ft pilot filter study, polymer doses increased the yield as shown in Figure 21 for 0.2:1 and 0.4:1 oxygen to primary sludge mixtures. In the centrifuge study, a mixture of 0.2:1 oxygen-waste-sludge to primary-sludge mixture was dewatered to 23.5% with polymer conditioning in a solid bowl centrifuge with a 6.0% feed concentration. The centrate contained 3,840 mg/1 of suspended solids.

Dewatering of oxygen sludges were tried on a two-cell one foot square Nichols filter press (Figure 22). The waste sludge conditioned with 4.7% ferric chloride and 14.4% lime by weight, dewatered to an average of 20.6% with a bulk cake density of 64 lb/ft^3. The mixture of sludge at 0.4:1 oxygen sludge to primary sludge, conditioned with 3.7% ferric chloride and 12.9% lime by weight, dewatered to 39.5% with a bulk cake density of 67.0 lb/ft^3.

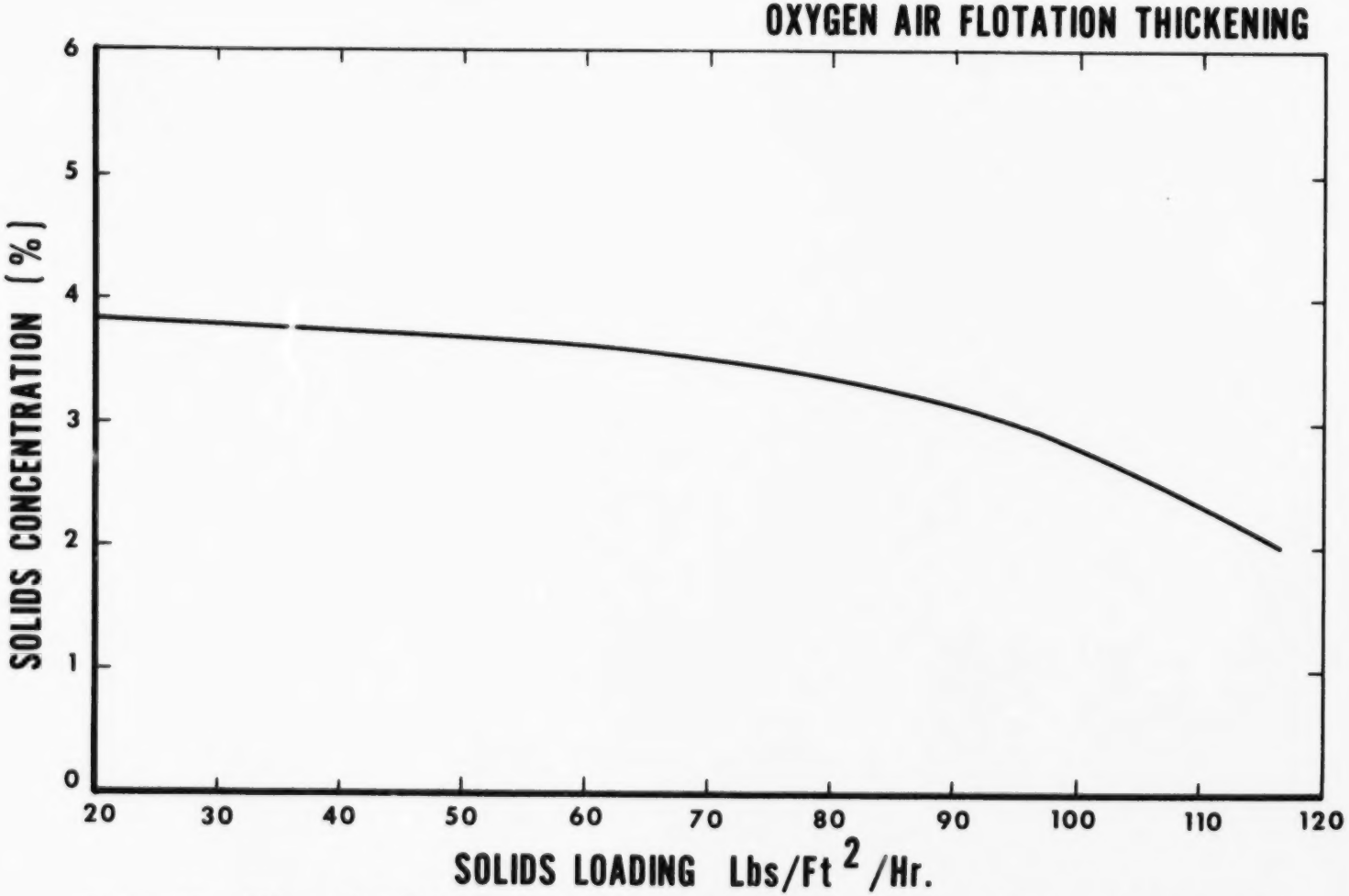

Figure 18. Effect of Solids Loading on the Concentration of Oxygen Sludge in Air Floatation Thickening

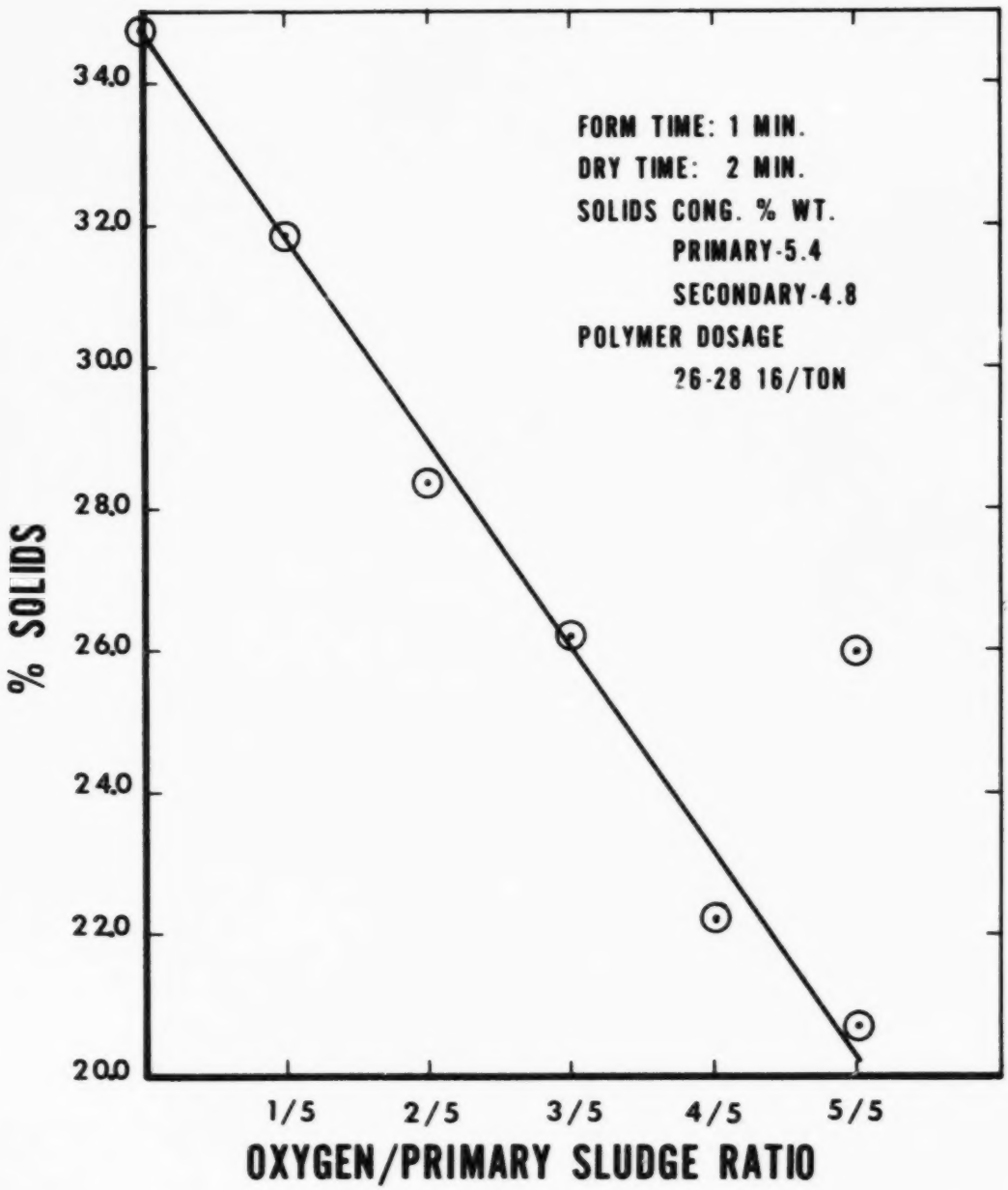

Figure 19. Effect of the Secondary to Primary Solids Ratio on Vacuum Filtration

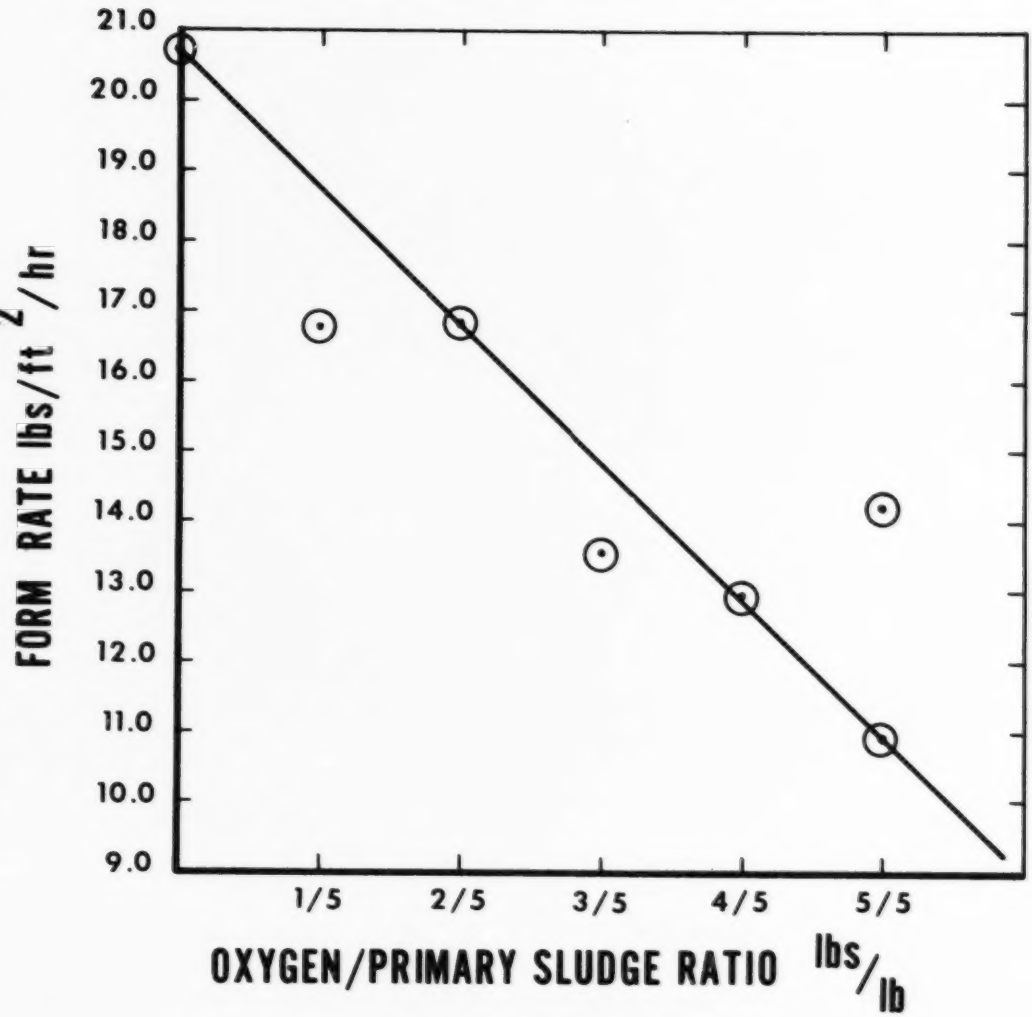

Figure 20. Effect of the Secondary to Primary Solids Ratio on Form Rate

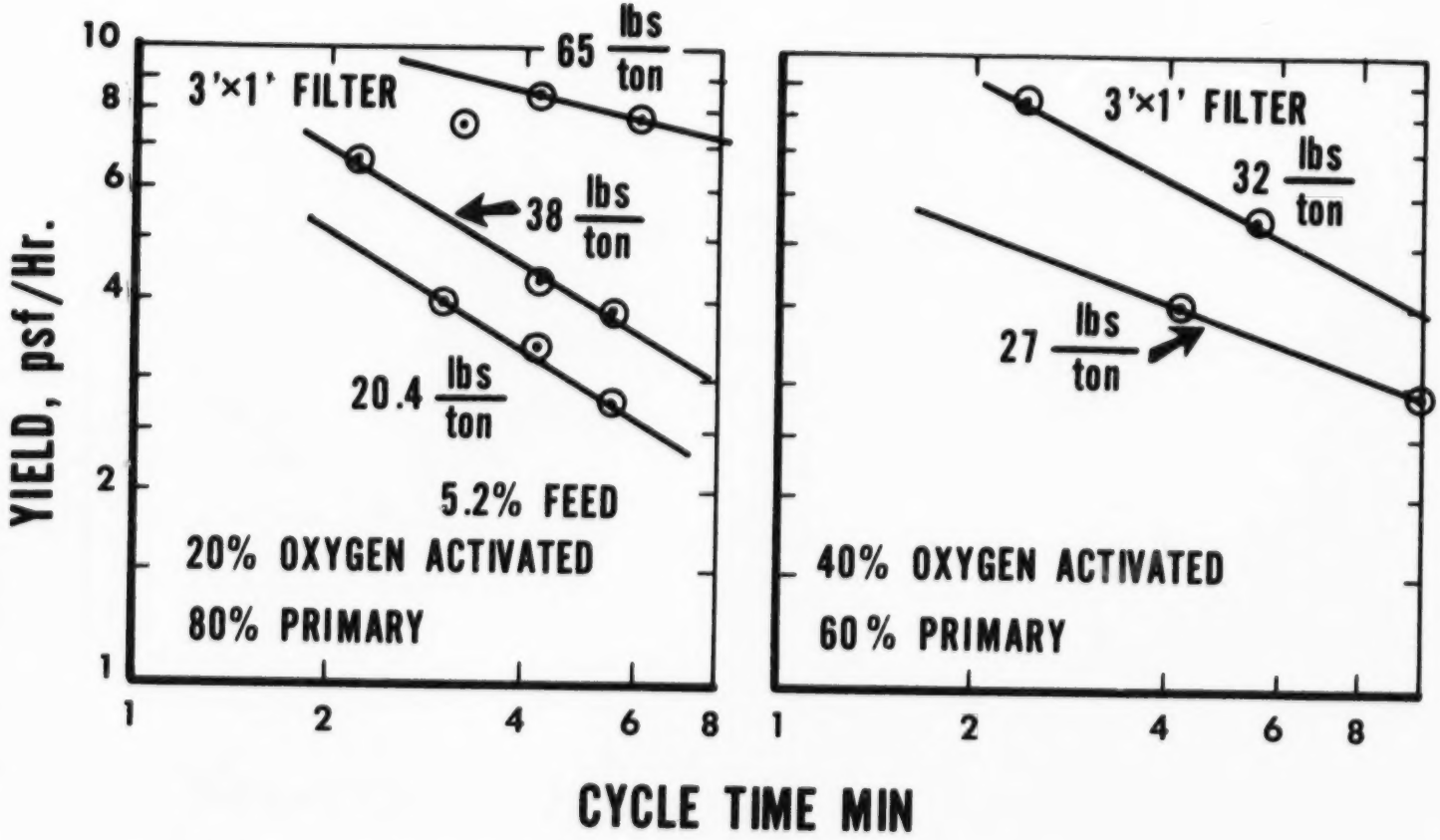

Figure 21. Effect of Polymer Dosage on Sludge Yield

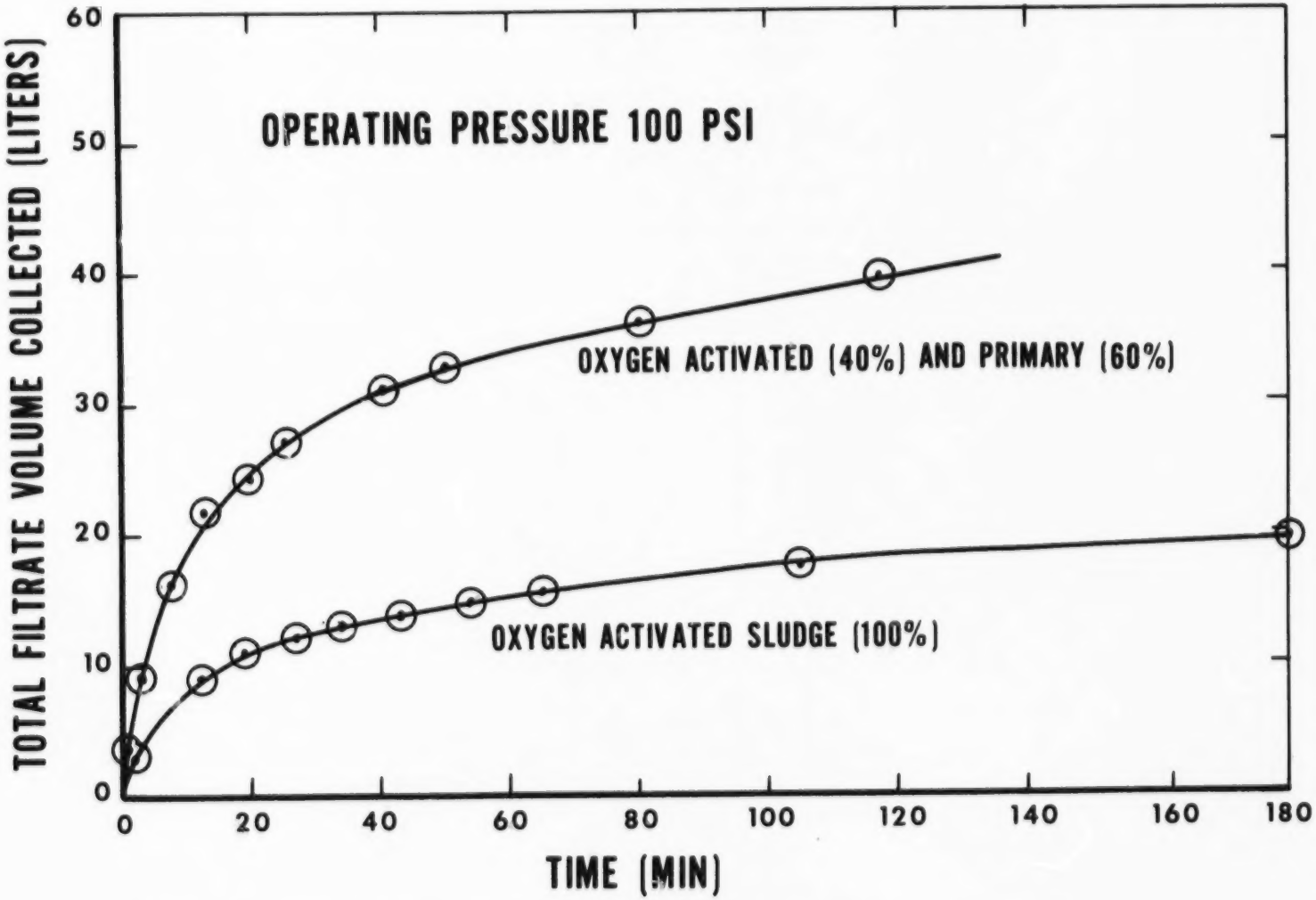

Figure 22. Oxygen Sludge Filtrate from a Filter Press as a Function of Time

SECTION VIII

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SECTION IX

PUBLICATIONS AND PATENTS

Presented at the 45th Annual Conference of the Water Pollution Control Federation, Atlanta, Georgia, October 1972.

No patents have been issued or requested as a result of the efforts of this study.